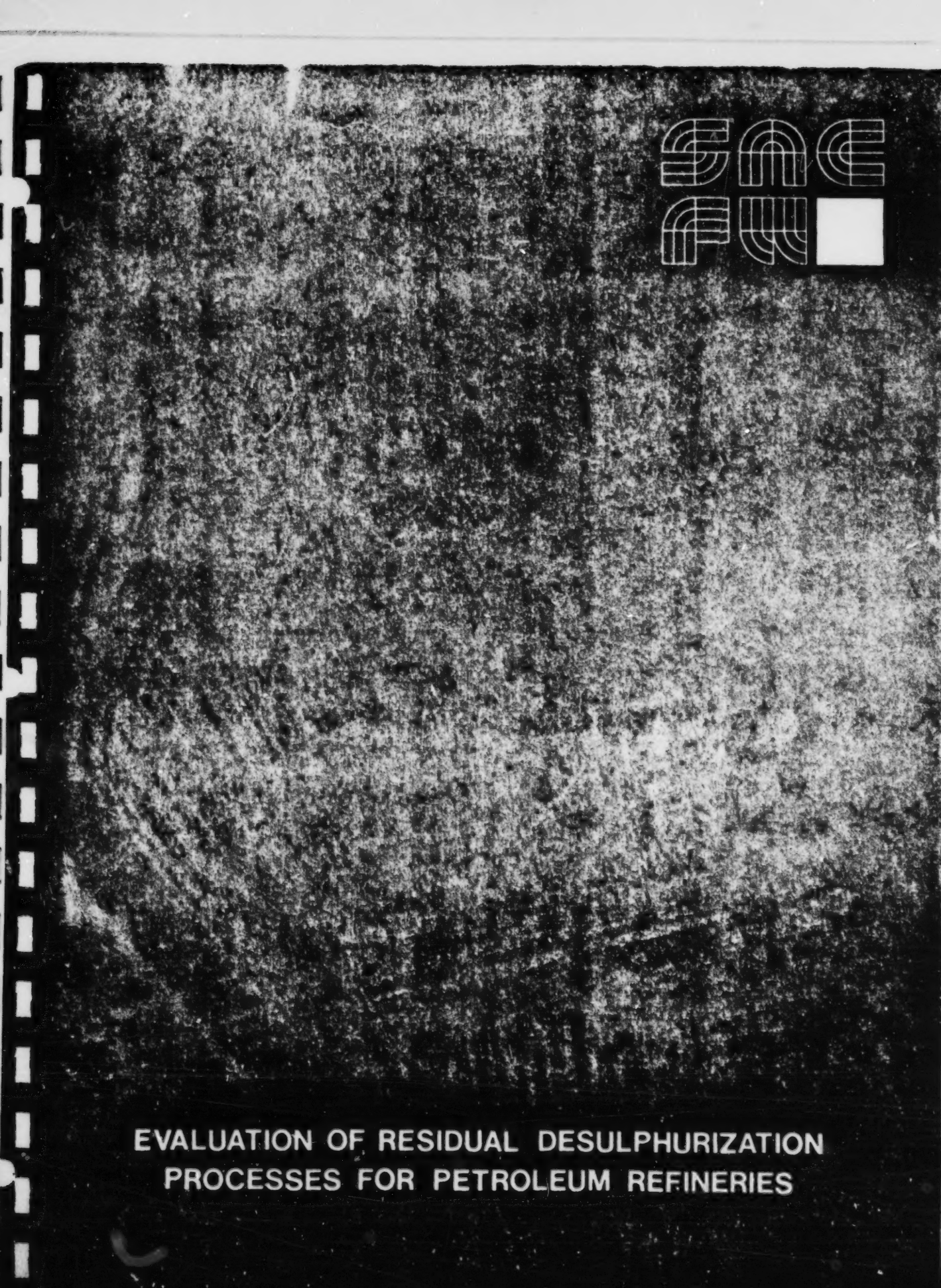
EVALUATION OF RESIDUAL DESULPHURIZATION
PROCESSES FOR PETROLEUM REFINERIES

EVALUATION OF RESIDUAL DESULPHURIZATION PROCESSES

FOR

PETROLEUM REFINERIES

BY

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ABSTRACT

SNC/FW Ltd. was retained by the Department of Supply and Services, Ottawa on behalf of the Department of Fisheries and Environment, Fuels Division to prepare a study presenting a technical and economic evaluation of the promising commercial processes for hydrodesulphurization of typical vacuum residua currently processed in Canada. The findings of the report would be utilized in the Department's current programme on abatement of sulphur dioxide emissions.

This study selected the residua form Tia Juana (Venezuela), the Agha Jari (Iranian Light) and the Canadian Light Sour Blend (S.E. Saskatchewan and Manitoba) crudes as representative of the high sulphur crude oils now in use in Canada.

The processes considered incorporate both fixed bed and moving bed hydrodesulphurization reactors. The facilities include a hydrogen plant using naphtha and light hydrocarbon byproducts from the hydrodesulphurizer as well as a gas treater to remove hydrogen sulphide, a sour water treater for environmental control, and offsites storage tanks.

Material balances are presented for each of the process facilities and related feedstock residua. Investment costs have been derived using Edmonton, Alberta as the plant location. Capital and operating costs are presented for each type of crude at two levels of desulphurization (0.5 and 1.0 weight % sulphur in the product boiling above 375°F) for residue charge rates of up to 30,000 BPSD.

The production costs for a barrel of desulphurized fuel oil for each case are presented taking into consideration credits for byproduct naphtha and sulphur and the cost of processing hydrogen sulphide through a Claus unit.

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1. INTRODUCTION

Today, most of world's crude oils contain a high amount of sulphur compounds. A large portion of these sulphur compounds is present as high molecular weight organometallics and asphaltenes which have high boiling points. These concentrate in the residue produced in the crude distillation unit. The sulphur content of the residue depends on the crude source, the sulphur distribution through the fractions, the types of high boiling sulphur compounds and the cut point of the residue. An overview of the sulphur contents of the crude oils throughout the world is presented in Table 1.1.

The residue can be used as a fuel within the refinery with the viscosity maintained by heating to proper temperatures. Alternately it can be blended with lighter cutterstock gas oils to meet viscosity requirements and sold as fuel oil product (no. 4, 5 or 6 grade). In either case, the combustion of these fuels for industrial, commercial and residential requirements generates a major source of sulphurdioxide emission. It is estimated that about half of the sulphur contained in the crude oils processed in Canada are emitted to the environment in this manner.

The development of technology for the desulphurization of heavy residua began in the 50's and attracted considerable attention in the 60's with mounting concern posed by stricter environmental regulations. The treatment of residua has to be performed under severe operating conditions of pressures and temperatures, more demanding than those for gas oils and distillates. The high metals content - primarily nickel and vanadium - causes rapid plugging and deactivation of the catalysts. Development activities were directed towards the development of catalysts that were resistant to these feed contaminants and operate under mild enough conditions so that reasonably long and stable operation could be maintained and be cheap enough for the spend catalyst to be discarded. These objectives have largely been met by a number of process licensors, and proven on a commercial scale.

Alternate processing routes to low sulphur fuel oil production are available although with limited desulphurization capabilities at times. Some of these indirect routes (1, 2, 3) include:

- Desulphurize the vacuum gas oil fraction (700-1100°F) to 85-90% desulphurization levels and blend with the vacuum residue (1100°F plus) (Figure 1.1).

- Solvent deasphalt the vacuum residue and desulphurize the deasphalted oil and blend with the desulphurized vacuum gas oil and straight run vacuum residue and or asphalt from the deasphalting unit (Figure 1.2).

There are currently a number of units in operation while additional units have been scheduled. The facilities at Kuwait, Shuaiba have been operating satisfactorily since 1967. Table 4.1.2 presents a survey of operating and planned units throughout the world.

1. INTRODUCTION (cont'd)

The main advantage of these schemes employing "indirect" desulphurization is the ease of desulphurization of the vacuum gas oil and deasphalted oil compared to that of residua. However, they cannot achieve the overall level of desulphurization that is possible with the direct desulphurization route.

Direct desulphurization can also be applied in combination with conventional visbreaking to produce premium distillates. It may also be attractive to employ direct desulphurization for preparation of Fluid Catalytic Cracking Unit chargestocks. An approach combining the residue desulphurization with conventional upgrading processes could improve the economics of refinery operation (4).

This study focuses on the direct desulphurization of typical vacuum residua processed in Canada.

TABLE 1.1

AVERAGE SULPHUR CONTENT OF WORLD CRUDE OILS

	Wt. %S
Saskatchewan - Fosterton Field (22.5°API)	2.90
Kuwait	2.53
Middle East (includes Egypt)	2.03
Middle East (excludes Kuwait)	1.77
West Texas (below 36° API)	1.77
Venezuela	1.70
Mississippi (Low Gravity)	1.60
West Texas Field	1.38
Alberta - Fennbig Valley (33.4° API)	1.09
California	1.00
U.S.A. Average	0.75
Alberta-Excelsior (36°API)	0.64
Canada (20-45° API)	0.44
East Texas Field	0.26
Alberta - Golden Spike (35° API)	0.23
North Africa	0.18
Alberta - Wizard Lake (37° API)	0.11
Far East	0.09

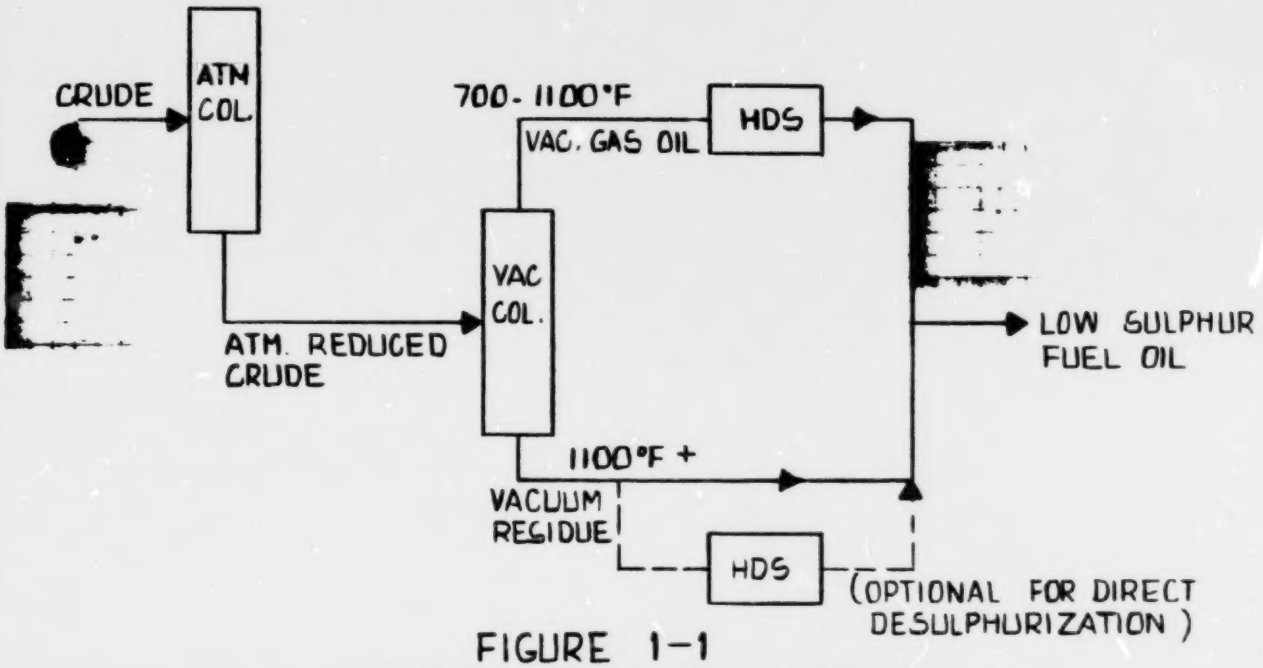

FIGURE 1-1
INDIRECT DESULPHURIZATION

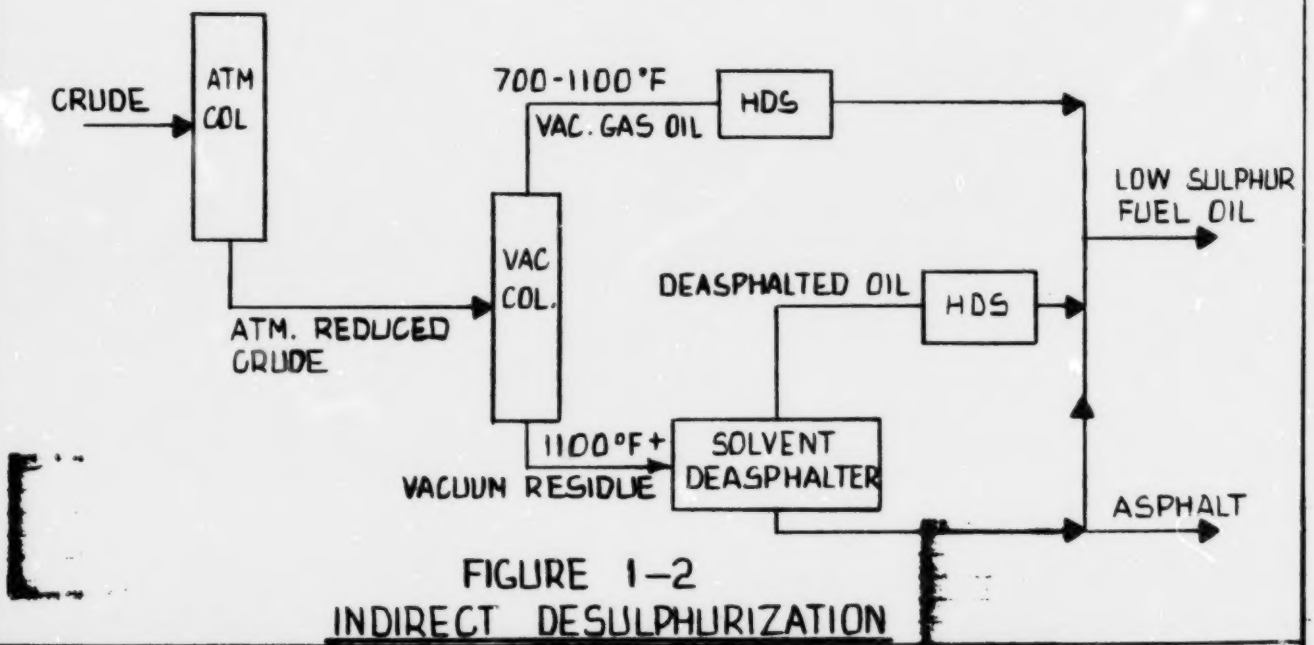

FIGURE 1-2
INDIRECT DESULPHURIZATION

CLIENT: ENVIRONMENT CANADA	FAIT/MADE: N. M.	VERIFIE/CHKD: PSK	DATE	
PROJ: EVALUATION OF DESULPH. PROCESSES	APPR.	ECH./SCALE: —	DATE REV	
INDIRECT DESULPHURIZATION	CONTR. / SUBDIV.	ELEMENT	NO	REV

FORMAT "A"
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2. BASIS OF STUDY

2.1 Feedstock

The three crudes have been selected covering the Canadian operation for this study:

- Tia Juana (Venezuela)
- Agha Jari (Iranian Light)
- Light Sour Blend (S.E. Saskatchewan and Manitoba)

Substantial amounts of sour mixed blend crude are processed in Canada. However, its sulphur content is very low and the residue does not require direct desulphurization.

Vacuum instead of atmospheric residues were chosen on the basis that a majority of Canadian refineries already include a vacuum tower. The vacuum residue desulphurization also enables the refiner to look at combining with the more common gas oil desulphurization or coking facilities to meet low sulphur fuel pool and increase the yield of premium products thus improving the overall economics.

These three crudes represent a cross-section of the ones processed in Canada. In direct residue desulphurization process, the amount of metals (nickel and vanadium) is of importance (5). Large amounts of metals cause rapid deterioration of the common cobalt molybdenum catalysts and require severe treating conditions in the reactor. The Tia Juana residue with over 700 ppm metals is one of the most difficult feedstocks to process. The Canadian Light Sour Blend residue has 80 ppm metals and represents one of the relatively easy feedstocks to desulphurize. In the middle of the spectrum is the Iranian Light residue with 250 ppm metals. Most of today's technology can economically handle chargestocks with metals up to 250 ppm. Thus, the selection of these crude types was based on considerations of investigating the ability of the processes to handle a range of residua on the relative difficulty of desulphurization and analyze their relative economics.

The properties of the residua were provided by two major Canadian oil refiners, the Department of Energy, Mines and Resources and the Bureau of Mines of the U.S.A. They were crosschecked with inhouse crude assay data and by the process licensors. The residue properties assumed for this study are summarized in Table 2.1.

2.2 Processes Evaluated

The following processes were chosen on the basis of proven commercial operation to date. They cover a spectrum of the fixed bed and ebullated bed desulphurization reactors:

Fixed Bed - UOP RCD-Unibon
Chevron VRDS
Gulf HDS
Exxon Residfiner

Ebullated Bed - HRI H-Oil

2. BASIS OF STUDY (cont'd)

2.3 Plant Configuration

The overall plant configuration scheme is presented in Section 5.

One of the major operating cost elements in direct desulphurization is the fresh hydrogen make-up required to satisfy the chemical consumption and compensate for the solution, purge and physical losses. A recent study [6] has shown that the purity of the hydrogen make-up could have a bearing on the overall economics of an integrated desulphurization facility such as the one considered here. The hydrogen purity of 97 mole % and the delivery pressure of 270 psig assumed in this study are close to the conditions suggested in the referenced study. The overall plant configuration assumes the maximum utilization of sour light gases generated as a byproduct in the desulphurization process as the hydrogen plant feedstock after sweetening in a gas treating unit with monoethanolamine (MEA). This scheme facilitates the use of light gas stream rich in hydrogen as the hydrogen plant feedstock rather than being assigned a lower value as refinery fuel. The sweetened light gas stream is supplemented by naphtha (C_5-375°F), which is also a byproduct in the desulphurization process, to satisfy the overall material balance. The typical range of naphtha content in the hydrogen plant feed is 30-40%. The effect of this assumption is the higher overall cost of hydrogen because the capital cost of the hydrogen plant is about 20% higher with naphtha feed than refinery or natural gas feed for equivalent hydrogen production capacity.

The sour water streams produced are combined and routed to a sour water stripper unit. The flash sour gases are sent to the gas treating unit while the stripped sour gases to the Claus Unit. The stripped water is reused in the desulphurizer and a small stream bled off for possible reuse in the crude desalter. The H_2S rich stream from the gas treating unit is assumed processed through Claus unit and a suitable tailgas clean-up unit to achieve a minimum of 99% sulphur recovery. The environmental quality standards for petroleum refineries [11] have been met by the inclusion of these additional facilities.

2.4 Estimation of Installed Costs

The installed "Inside the Battery Limit" costs for U.S.A. Gulf Coast location were supplied by the process licensors. Independent material balances were derived utilizing published data as well as yield information supplied by the licensors enabling to estimate the major equipment sizes and derive installed costs for Edmonton, Alberta location. This procedure brought to light the adjustments required to the U.S. Gulf Coast investment costs to account for services facilities which have not been included in the licensors' estimates as well as higher costs for the Alberta location due to additional costs for transportation and lower labour productivity. A detailed discussion on the approach to capital cost estimating is presented in Appendix C.

2. BASIS OF STUDY (cont'd)

2.4 Estimation of Installed Costs (cont'd)

The principal difference amongst the technologies lie in the reaction system and the catalysts employed. However the product flash recovery and recycle hydrogen treatment sections have been assumed to be identical for all fixed bed processes.

The recycle hydrogen scrubbers have been omitted in the case of H-Oil.

2.5 Process Economics

Appendix A presents capital charges and unit costs for labour and utilities assumed in the economic analysis. The unit costs presented for steam, cooling water and Boiler Feed Water include the fixed capital charges and the variable costs.

Two types of financing -utility and equity financing- have been assumed. In the case of utility financing the interest rate is 8% over a mortgage period of 20 years yielding an overall loan rate of 10.3%. In the case of equity financing plant depreciation rates of 6 and 4% for inside the battery limit and outside the battery limit investment have been assumed with the debt interest of 15%.

These two types of financing represent the extremes. The debt/equity ratio employed would in reality determine the finance charges.

TABLE 2.1

FEEDSTOCK PROPERTIES

	Tia Juana	Iranian Light Agha Jari	Canadian Light Sour Blend
Cut Point	1020°F+	1050°F+	1050°F+
°API	6.5	7.6	7.1
Conradson C.R.	20.0	14.1	21.5
Sulphur Wt.%	3.14	3.04	3.40
Nitrogen Wt.%	0.70	0.59	1.09
Metals (Ni + V) ppm	705	250	82

3. OPERATING CONDITIONS AND VARIABLES

3.1 Chemistry

The selection of catalyst, operating temperature, operating pressure, space velocity and hydrogen to feed ratio are major process considerations for the desulphurization processes. The proper combination of these variables is necessary to obtain the desired degree of selective hydrogenation of sulphur compounds. Side reactions resulting in the hydrogenation of nitrogen, oxygen, organic metals and unsaturated hydrocarbon compounds contained in the Crude residue also occur. All of these reactions help improve the product quality.

Sulphur compounds such as organic sulphides and disulphides, thiophanes, tiophenes and traces of mercaptans are present in the unit feed. The relative amount of these sulphur compounds depends upon the particular petroleum fraction, source of crude and prior processing.

On the basis of studies on hydrogenating pure sulphur compounds, it can be concluded that the primary reaction is the breaking of the carbon-sulphur bond and the subsequent reaction of hydrogen with the sulphur atom producing hydrogen sulphide. The mercaptans, thioethers and disulphides are readily hydrogenated and with carbon atom saturating the hydrocarbons at fairly mild conditions of temperature and pressure.

When cyclic sulphur compounds are present, such as thiophenes, the compound is first converted to thiophane by saturating the olefin bond followed by breakage of the ring and formation of the parent hydrocarbon plus hydrogen sulphide. Ring compounds are generally more difficult to hydrogenate and the aromatic ring is not saturated or broken except under severe hydrogenation conditions.

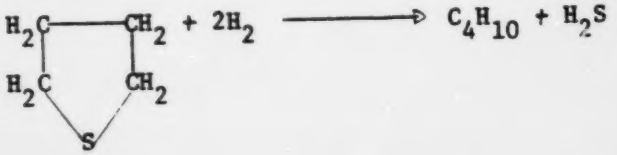
Although the exact composition of some of the sulphur compounds in the feed is not known, typical hydrogenation reactions of known sulphur compounds are shown as follows:

Mercaptans $RSH + H_2 \longrightarrow RH + H_2S$

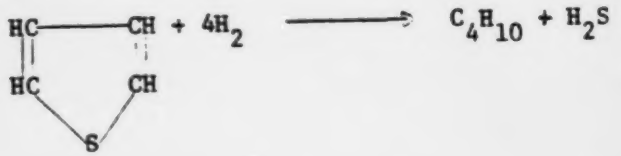
Thioether $RSR' + 2H_2 \longrightarrow RH + R'H + H_2S$

Disulphides $RSSR' + 3H_2 \longrightarrow RH + R'H + 2H_2S$

Thiophanes



Thiophenes

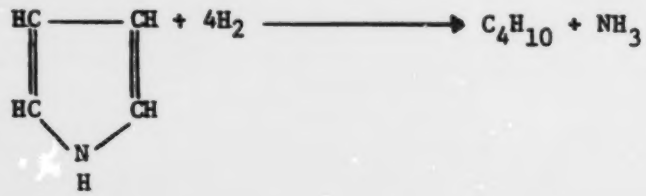


3. OPERATING CONDITIONS AND VARIABLES (cont'd)

3.1 Chemistry (cont'd)

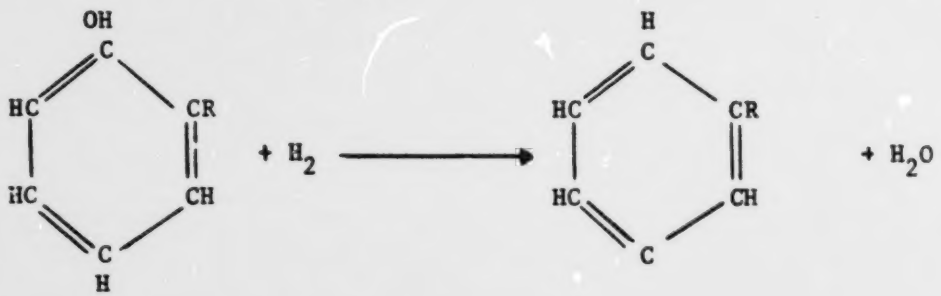
The hydrogenation of nitrogen compounds in the feed results in the liberation of ammonia. Here again, the relatively unknown composition of the reactants makes it difficult to indicate the precise mechanism of nitrogen removal. One reaction that is suggested is the hydrogenation of pyrrole, a nitrogen compound similar to thiophene.

Pyrrole



The hydrogenation of oxygen compounds is of less importance than the reactions with sulphur and nitrogen compounds. This is because of the negligible amounts present in the crude residue. However, removal of the oxygen compounds help to produce a more stable product. The following reaction shows the typical hydrogenation of a high boiling phenol derivative.

Phenol Derivative



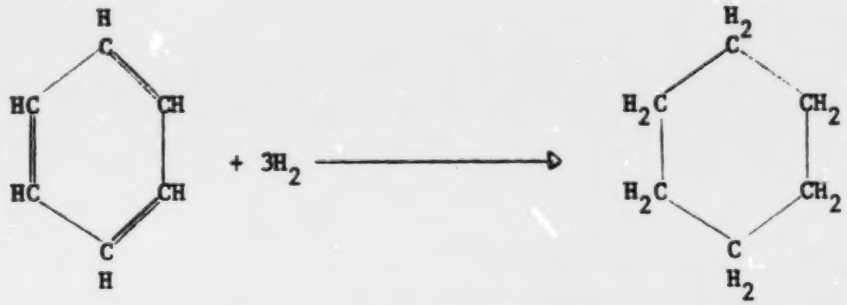
The exact reactions involving organometallic compounds are even harder to define than the hydrogenation of oxygen and nitrogen compounds. However, it is suspected that metals are removed by hydrogenation of the organometallic compounds and that the metals are retained on the catalyst. High metal contents in the feed are undesirable because the deposit of metals normally deactivates the catalyst.

The hydrogenation of unsaturated hydrocarbons is common in all hydrodesulphurization units. The degree of this reaction normally depends upon the severity of the operation. A common hydrogenation reaction is the saturation of aromatic compounds. The hydrogenation of benzene to cyclohexane is shown below.

3. OPERATING CONDITIONS AND VARIABLES (cont'd)

3.1 Chemistry (cont'd)

Benzene



Because of the types of hydrogenation reactions taking place, the resulting hydrocarbon compounds are of lower density than the original compounds. For this reason, the liquid volumetric yield is usually more than 100% based on the charge. However, stripout of the light components normally reduces this yield of finished product to just under 100%.

3.2 Quality of Feedstock

The quality of residue varies considerably depending on the origin and gravity of the crude from which it was produced.

The feed residue properties which have an important influence on the difficulty and cost of residual oil desulphurization are: sulphur level, metal contents, conradson carbon or asphaltene content, nitrogen content, and oil viscosity. Oils of higher sulphur content require more hydrogen for desulphurization. The metals content and the asphaltene content both have a deteriorating effect on catalyst life; Vanadium and nickel deposit in the upper part of the bed deactivating that portion of the bed insofar as sulphur removal is concerned. The deposited metals also tend to catalyze the deposition of more metals and hence the metals content builds to a high level. The asphaltenes apart from being relatively resistent to desulphurization can lead to coke deposition, which in turn affects the activity and life of catalyst. The evolution of large quantities of ammonia produced from nitrogen content interferes with hydrogenation on active catalyst sites, and the diffusion of reactants into the catalyst diminishes as viscosity increases.

3.3 Process Variables

The hydrodesulphurization processes for residua are designed to provide low sulphur fuel oil by hydrogenation of the feedstock.

Various degrees of hydrogenation can be obtained depending on the operating conditions employed. The following basic variables affect the degree of hydrogenation:

3. OPERATING CONDITIONS AND VARIABLES (cont'd)

3.3 Process Variables (cont'd)

- Space velocity
- Hydrogen partial pressure
- Reactor temperature

3.3.1 Space Velocity

Space velocity is a measure of the contact time between hydrocarbon and catalyst. It is defined as the volume per hour of hydrocarbon feed divided by the catalyst volume and can be expressed as follows:

$$V/H/V = \frac{\text{Volume of Residuum Feed/Hr}}{\text{Volume of Catalyst}} = \frac{\text{Cubic Ft/Hr}}{\text{Cubic Ft}}$$

Reducing the space velocity tends to increase the degree of hydrogenation. However, since the volume of catalyst in the system is fixed, space velocity can only be reduced by decreasing the feed rate.

3.3.2 Hydrogen Partial Pressure

Hydrogen partial pressure is defined as that part of the total vapor pressure contributed by the hydrogen present in the reactor system. Increasing the hydrogen partial pressure results in an increase in the degree of desulphurization. Also, increasing the hydrogen partial pressure tends to slow the rate of catalyst deactivation and results in longer catalyst life. The other effect of increasing hydrogen partial pressure is the increase in hydrogen consumption, because of increased saturation of olefins and aromatic hydrocarbons.

The hydrogen partial pressure can be controlled by the following operating parameters:

- Increasing the system pressure increases the hydrogen partial pressure.

- Increasing the hydrogen concentration by removing Ammonia and hydrogen sulphide from the recycle gas increases the hydrogen partial pressure. The hydrogen purity is also controlled by the amount of makeup hydrogen added to the unit (which affects the hydrogen purged from the system).

- Increasing the recycle gas to hydrocarbon feed ratio also increases the hydrogen partial pressure.

3. OPERATING CONDITIONS AND VARIABLES (cont'd)

3.3.3 Reactor Temperature

Reactor temperature is the major variable used to control the level of desulphurization. Hydrogenation (desulphurization) increases with temperature with the reaction rate doubling about every 60°F. As the catalyst deactivates, the reactor temperature is gradually increased to maintain the same level of desulphurization. The run is usually terminated at a temperature at which increasing hydrocracking results in a sharp upturn in hydrogen consumption. Increased temperature may be used to offset lower space velocity but generally increased temperature results in increased demetallization as well as increased desulphurization and promotes higher rate of catalyst coking.

Summarizing the effects of the previously mentioned variables, an increase in hydrodesulphurization can be obtained by the following:

- Decreasing the space velocity by decreasing feed rate.
- Increasing the reactor pressure.
- Increasing the recycle gas hydrogen purity.
- Increasing the recycle gas rate.
- Increasing the reactor temperature.

All of the above variables can affect product quality. Normally, as the catalyst deactivates from start of run to end of run, the reactor temperature is increased to maintain the required level of desulphurization.

3.4 Catalyst Characteristics

From the extensive research and patent activity on catalyst preparation and testing, certain generalizations may be drawn (7). It has been suggested that a desirable catalyst have at least 250 m^2/g of surface area with pore volume of at least 0.6 cc/g and with pore diameter of 50 to 90 A. The desirable combination is either cobalt or nickel with molybdenum on an alumina carrier. Since silica promotes hydrocracking, the suggested maximum silica content is 2.0%. However, it is important to note that the catalyst properties vary widely from the "norms" given above and there is also variation in their desulphurization activity.

Catalyst deactivation rate can be reduced by the following:

- Eliminating feed contaminants (double desalting of crude and filtering residue).

- Maximizing hydrogen partial pressure.

- Minimizing reactor temperatures.

- Running clean feeds with qualities favoring mild reactor conditions.

4. REVIEW OF COMMERCIAL PROCESSES

4.1 General Overview

A number of commercial processes are available for desulphurization of vacuum residua. They can be broadly classified as:

- fixed bed processes
- "moving" bed processes.

Among the fixed bed processes are those offered by UOP (RCD Unibon), Chevron (VRDS), Gulf (HDS), Exxon (Residfining), British Petroleum and American Oil Company. In general they can economically process residues containing not more than 200 ppm metals.

All these processes employ fixed catalytic bed reactor systems in which the feed is contacted with hydrogen at elevated hydrogen partial pressures (typical total operating pressures range from 1500 to 3000 psig) and temperatures in the range of 650 - 800°F (8). The major difference between these systems is the catalyst employed in the reactors and its ability to achieve the desired degree of desulphurization at favourable space velocities to minimize reactor volume, minimize coking and hydrocracking reactions to reduce chemical hydrogen consumption and increase the yield of valuable products and sustain stable operation for extended periods of time.

The degree of desulphurization can be varied by basically changing the reactor system design. It is possible to initially design the product recovery section for higher levels of desulphurization with flexibility provided in the reaction system for future additions. For example Gulf offers different designs tailored to meet the sulphur content of the fuel oil product. Depending on the sulphur content of the feedstock, the Gulf Type II process can reduce the sulphur content in the fuel oil to about 1.0 wt.%. Type III and Type IV processes are offered in case the sulphur content has to be reduced to 0.3 and 0.1 wt.% respectively. This is facilitated by employing the new coke resistant catalysts instead of the conventional types used in the Type I (predecessor of Type II) and II processes. There is some flexibility for processing a variety of residua and to meet the changing requirements of sulphur levels in the product by converting from one type to another. For example, the low cost Type II system having capability to 1.0 wt.% sulphur could be initially selected, however, having provisions to conversion to Type III to achieve 0.3 wt.% sulphur. Figures 4.1 to 4.4 illustrate the basic differences in the features of Gulf designs (9).

The major processes offered utilizing "moving" bed reactors are:

- H-Oil, Hydrocarbon Research Inc.

- Shell Desulphurization, bunker flow reactor

4. REVIEW OF COMMERCIAL PROCESSES (cont'd)

4.1 General Overview (cont'd)

In contrast to the fixed bed systems which have to be shut down usually 6-12 months to discard the spent catalyst and replace with fresh catalyst, the "moving" bed systems feature continuous addition of fresh catalyst and withdrawal of spent catalyst. H-Oil have a patented ebullated bed reactor where the catalyst is kept in motion. The Shell process employs a bunker flow reactor wherein the catalyst travels down the length of the reactor. The advantages claimed for these processes are that the catalyst activity would be uniformly maintained resulting in uniform product distribution and purity and less intricate process control. The operating problems arising from bed channelling, pressure drop and uneven temperature profile across the reactor length are also overcome. A typical ebullated bed reactor of HRI design is shown in Figure 4.1.5.

It is of significant interest to note that of all the processes investigated in this study, only H-Oil could economically handle the Tia Juana vacuum residue. The H-Oil design differs from the others in that a vacuum tower is included to generate a 650°F+ gas oil stream from the reactor effluent which is recycled to mix with the reactor fresh feed to "improve the catalyst performance" (10). Two reactors in series are employed. The catalyst in the first reactor is primarily for demetallization while the second reactor catalyst primarily performs desulphurization. (This feature however is also offered in the fixed bed processes.) Acid gas treatment of the recycle gas is not employed and the hydrogen partial pressure is maintained at desired levels by increasing the total pressure as required.

4.2 Commercial Experience

The operating experience of various licensors of residue desulphurization processes around the world have been referred to in Section 1. Table 4.2.1 has been compiled from an in-house literature search and the information provided by the process licensors.

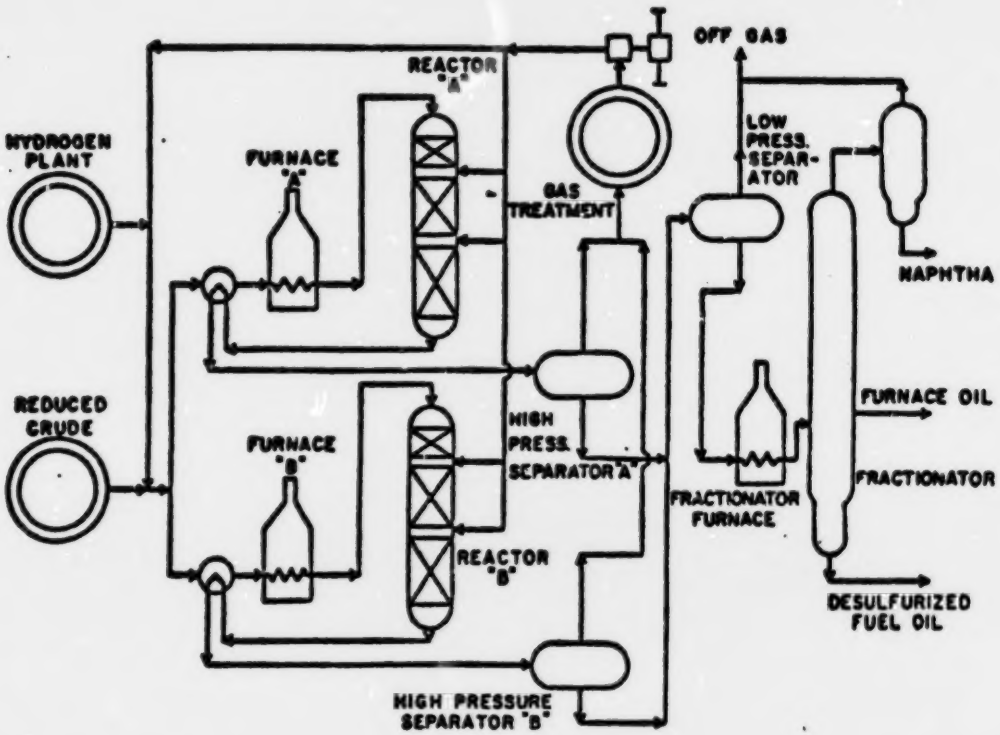


FIGURE 4.1.1
FLOW DIAGRAM OF COMMERCIAL HDS UNIT

GULF HDS PROCESS TYPE I

FIGURE 4.1.2

GULF HDS - TYPE II

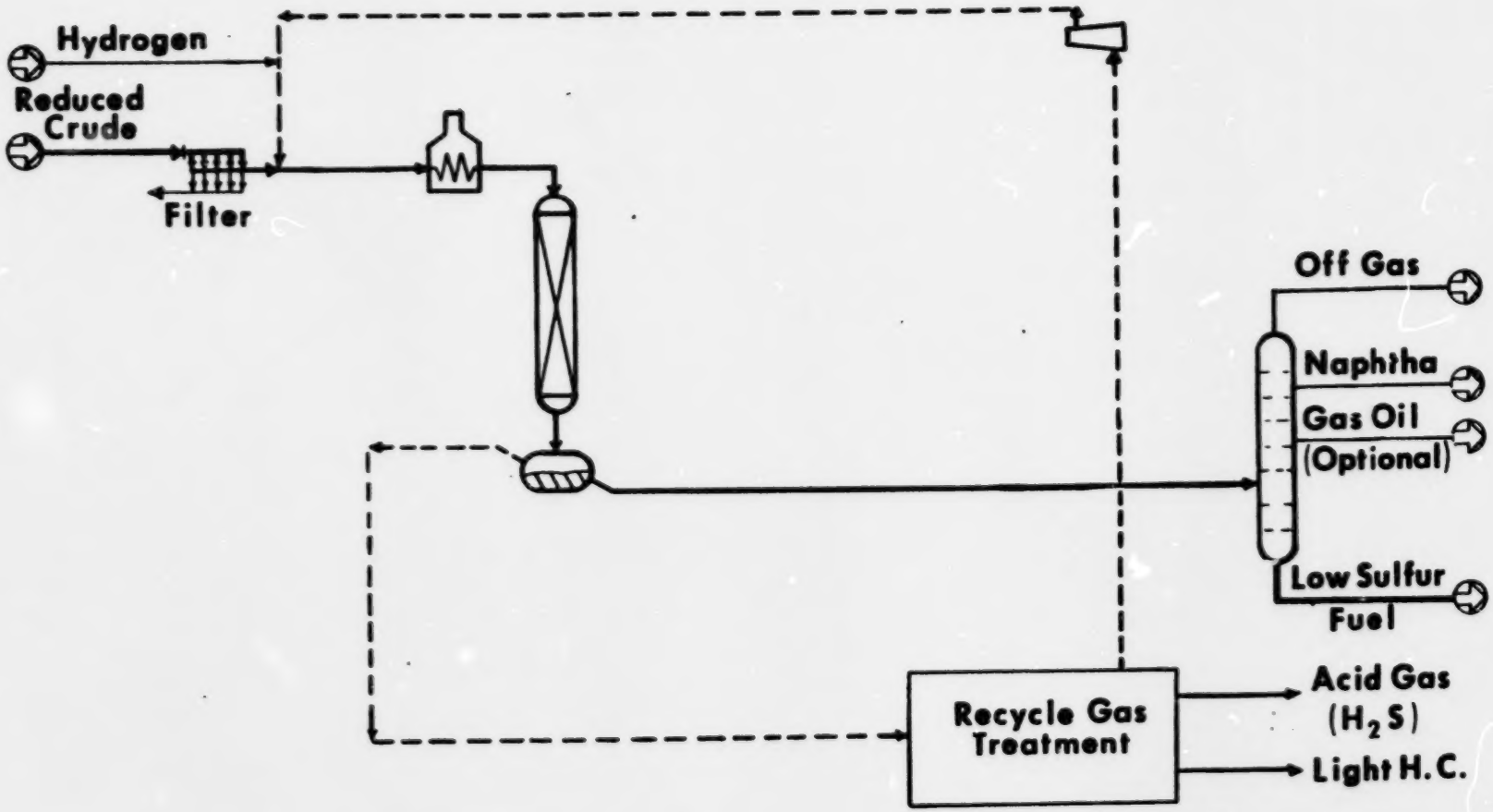


FIGURE 4.1.3

GULF HDS - TYPE III

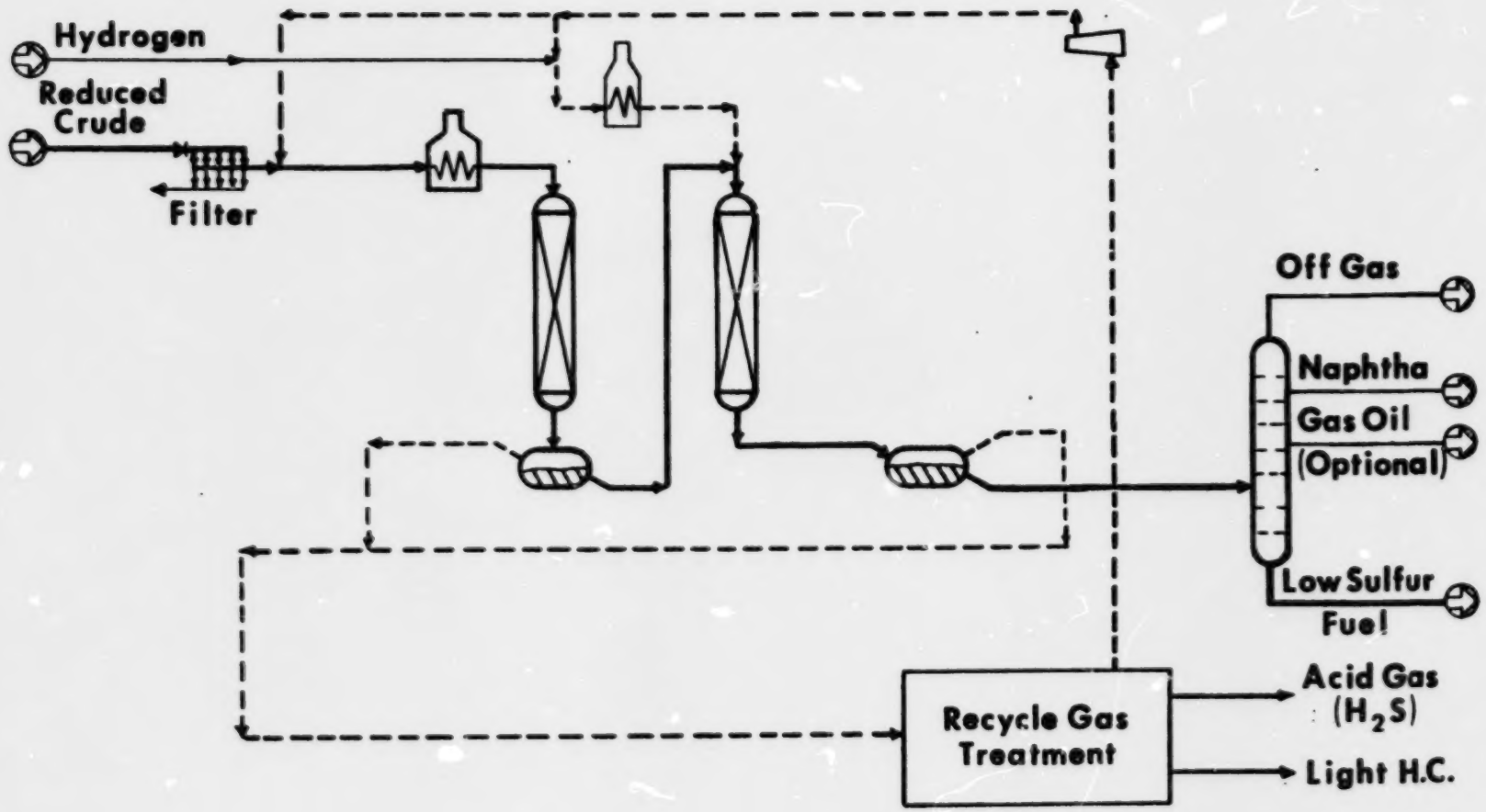
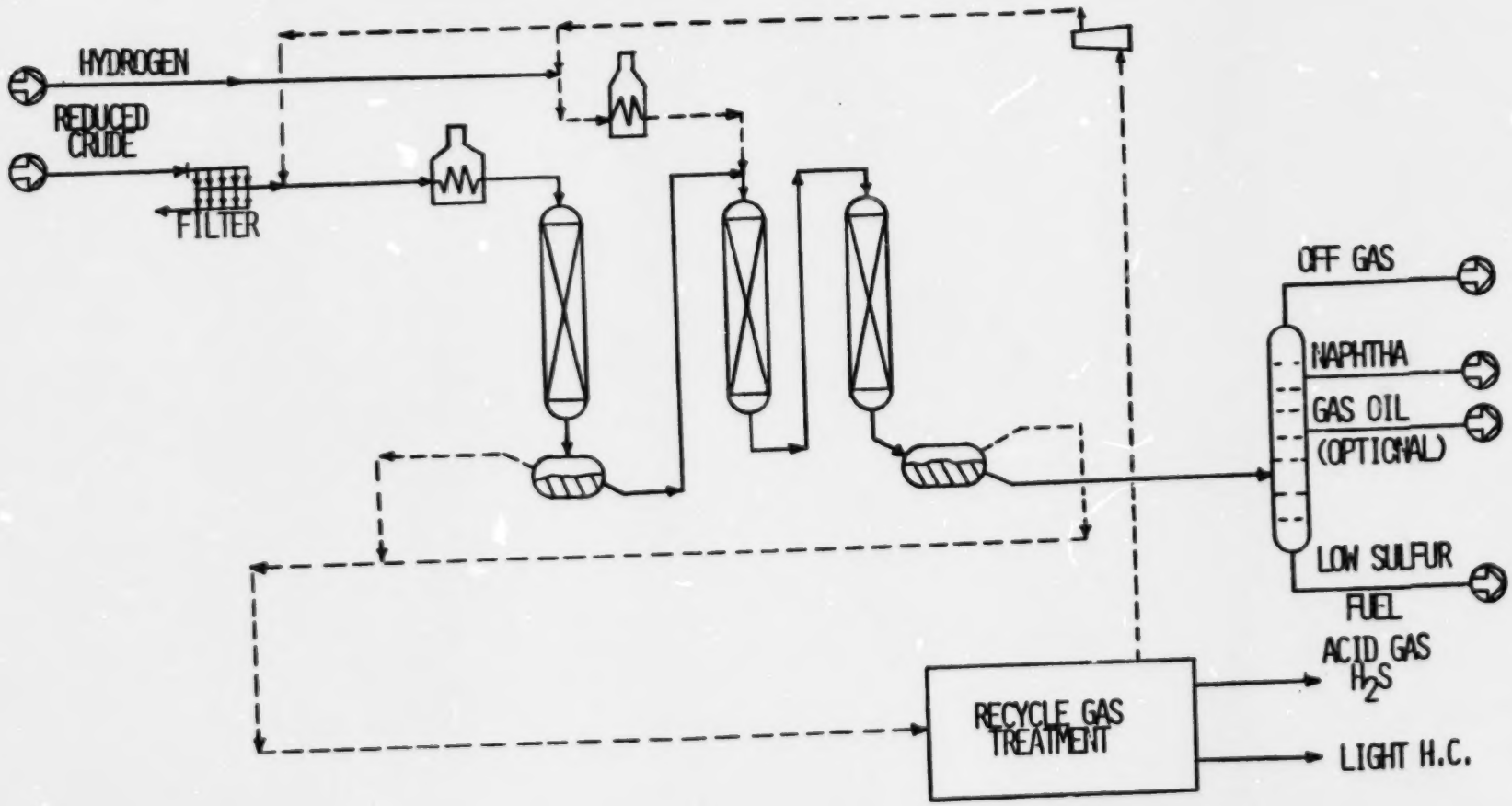


FIGURE 4.1.4

GULF HDS-TYPE IV



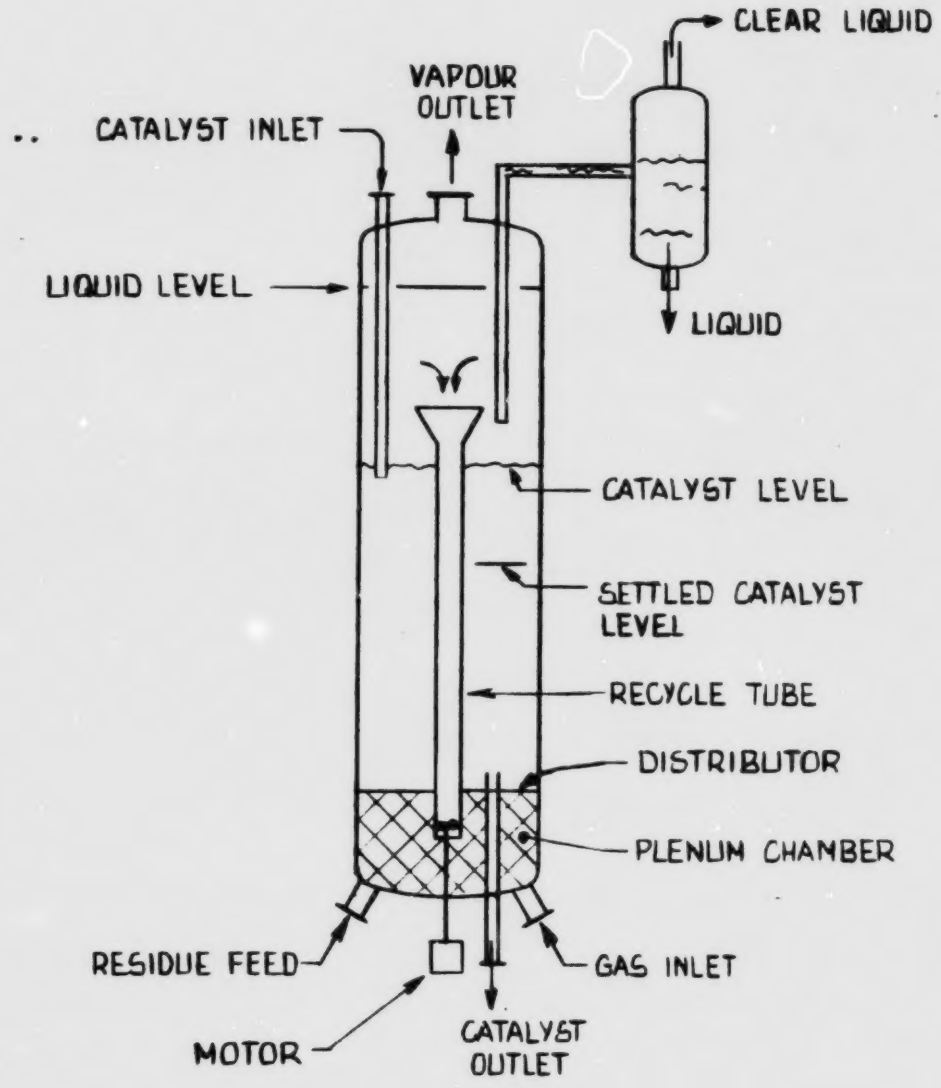


FIGURE 4-1-5

CLIENT ENVIRONMENT CANADA	FAIT/MADE N. M	VERIFIE/CHKD PSK	DATE	
PROJ. EVALUATION OF DESULPH. PROCESSES	APPR.	ECH./SCALE —	DATE REV	
H-OIL EBULLATED BED REACTOR	CONTR. / SUBDIV.	ELEMENT	NO	REV

FORMAT "A"
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TABLE 4.2.1

OPERATING RESIDUE DESULPHURIZATION UNITS

Process & Location	Start-up Date	Feed Type	Service	Design Capacity BPSD	Comments
Residfining (Exxon)					
Exxon, Baytown, Texas	1977 (1st Quarter)	ARC		60,000	
Unidentified	Under Construction	ARC		30,000	
Unidentified	Under Construction	ARC		N/A	
Unicracking/HDS (Union)					
Maruzen Oil, Chiba, Japan	June 1976	ARC		60,000 (30,000 now on stream)	.91 SF
H-Oil (Hydrocarbon Research)					
Exxon, Bayway, N.J.	1971			16,500	Destroyed by explosion after 14 weeks on-stream
Cities Service, Lk. Charles, La.	1963	ARC, HCO	Conversion	2,500	Expanded to next unit
Cities Service (Expanded Unit) (Lk. Charles, La.)	June 1972	VRC, PDA	Conversion & Desulphurization	6,000	Now being revamped to run solvent coal
KNPC, Shuaiba, Kuwait	October 1968	VRC	Conversion & Desulphurization	28,800	
Pemex, Salamanca, Mexico	1973	VRC, PDA Lube Ext.	Conversion & Desulphurization	18,500	
Shell Hydrodesulphurization					
Koppartrans Ref., Gothenburg, Sweden	March 1973	ARC	Desulphurization	3,000	Moving bed Demonstration Unit.
Seibu Sekiyu, Onada City, Japan	1976	ARC	Desulphurization	47,000	

TABLE 4.2.1 (cont'd)

OPERATING RESIDUE DESULPHURIZATION UNITS

Process & Location	Type	Start-up Date	Feed Type	Service	Design Capacity BPSD	Comments
Texfining (Texaco)						
Texaco, Wilmington, Ca.		June 1962	VRC	Conversion	12,000	Shutdown
Resid HDS (Gulf)						
Nippon Mining, Mizushima, Japan	I	Jan. 1970	ARC	Desulphurization	28,000	
Idemitsu Kosan, Himeji, Japan	II	Apr. 1972	ARC	Desulphurization	40,000	
Okinawa Sekiyu Seisei, Okin, Japan	II	May 1972	ARC	Desulphurization	38,000	
Mitsubishi Oil, Mizushima, Japan	II	May 1974	ARC	Desulphurization	45,000	
Idemitsu Kosan, Aichi, Japan	IV	Nov. 1975	ARC	Desulphurization	50,000	
VRDS Hydrotreating (Chevron)						
Chevron, El Segundo, Calif.		1976	VRC	Desulphurization	24,000	
RCD Isomax (UOP/Chevron)						
Idemitsu Kosan, Japan		Oct. 1967	ARC	Desulphurization	40,000	
Aminoil, Kuwait		Feb. 1969	ARC	Desulphurization	35,000	
Kashima Sekiyu, Japan		Aug. 1970	ARC	Desulphurization	45,000	
BOC Isomax (UOP/Chevron)						
Charter Oil, Houston, Texas (Formerly Signal Oil & Gas)		Fall 1968	VRC	Conversion	10,000	Shutdown
RCD Unibon (UOP)						
Unidentified		1975	ARC		Not Available	
				Total in operation	552,800	

Note:
HCO - Hvy. Cycle Oil; ARC-Atmos, Red. Crude; VRC - Vac. Red, Crude;
PDA - Propane Deasph. Btms; Lube Extr. - Lube Extracts

5. TECHNICAL/ECONOMIC ASSESSMENT

5.1 Process Design

5.1.1 Hydrodesulphurization (HDS) Unit

A typical process flow diagram for the HDS Unit is shown in Figure 5.1.1 Minor variations in the product flash and distillation sections may be employed to optimize the heat recovery for different types of feedstocks.

The vacuum residue or the Vacuum Tower Bottoms (VTB) is fed to the unit at 450°F via a surge drum. The residue is passed through a filter to remove particles that may plug the catalyst bed, mixed with hydrogen and brought to the reaction temperature in the fired heater H-1. The exothermic heat of reaction is removed by quenching with cold recycle hydrogen. (The reactor effluent liquid can also be recycled to the reactor inlet as the quench medium.) The reactor (R-1) products give up heat to the incoming feed and then are flashed in the Hot High Pressure Flash Drum (V-2) to separate the liquid and vapour streams. The hot vapours are cooled by heat exchange with internal streams, condensed in E10 and sent to the Cold High Pressure Drum (V-3). Process water is injected ahead of the heat exchanger train to prevent plugging by ammonium salts deposition.

The separated vapour stream from V-3 is scrubbed with fresh process water in the water scrubber (T-1) and with monoethanolamine (MEA) in MEA Scrubber (T-2) to remove ammonia and hydrogen sulphide formed in the reactor. The removal of these components as well as a small purge stream from the recycle gas is necessary to maintain adequate hydrogen partial pressure in the reactor. The recycle gas is then compressed, mixed with fresh 97 mole % hydrogen and routed to the reactor. The water stream from V-3 contains dissolved ammonia and hydrogen sulphide and is sent to the sour water stripper unit. The flash vapours from the Hot Feed Drum (V-5) are condensed, mixed with the hydrocarbon phase from V-3 and flashed in the Cold Low Pressure Drum (V-4) to remove sour light gases which are routed to the Gas Treating Unit. The combined liquid feed from V-5 is fractionated in T-3 to naphtha and desulphurized fuel oil products. The fractionator overhead gases are sent to the Gas Treating Unit for removal of hydrogen sulphide. A portion of the naphtha product is sent to the Hydrogen Plant to supplement the feedstock as required and the excess sent to storage. The desulphurized fuel oil product is cooled first to generate steam and then with air to a safe storage temperature.

Superheated steam is employed to strip the fuel oil product to meet the flash point requirements. This generates a sour water stream in the reflux drum (V-7) which is sent to the sour water stripper unit for treatment.

5. TECHNICAL/ECONOMIC ASSESSMENT (cont'd)

5.1.1 Hydrodesulphurization (HDS) Unit (cont'd)

Depending on plant size, the feedstock and the refinery utility balance it may be economically justifiable to include low pressure steam generators in the hot vapour circuits and power recovery turbines in the streams let down from high pressures.

5.1.2 Hydrogen Unit

Figure 5.1.2 represents a typical process scheme for the manufacture of hydrogen using mixed liquid naphtha-light hydrocarbon gas feed. The preferential feedstock is the hydrogen rich light hydrocarbon gases. Naphtha is used as the supplemental feedstock to satisfy the hydrogen plant requirements when necessary.

Naphtha from the HDS Unit stored in tank TK101 is vapourized and mixed with the light gases which had been preheated in the convection section of the reformer furnace. Trace quantities of sulphur are converted to hydrogen sulphide by reacting with hydrogen rich stream in the reactor (R101). Hydrogen Sulphide is absorbed by Zinc Oxide in R102. The sulphur free feed is mixed with superheated steam and passed through catalyst tubes in the furnace H102. The following reactions take place:

$$C_nH_{2n} + nH_2O \rightleftharpoons nCO + (2n + 1)H_2$$

$$C_nH_{2n} + 2nH_2O \rightleftharpoons nCO_2 + (3n + 1)H_2$$

The reactions are endothermic. The heat of reaction is provided by the furnace. The furnace effluent consisting of an equilibrium mixture of carbon monoxide (CO), carbon dioxide (CO_2), hydrogen (H_2) and methane (CH_4) is cooled and passed over the shift converter catalysts in R103A and B at successively lower temperatures wherein the following well known shift reaction takes place:

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

The CO_2 is absorbed by a circulating solution of potassium carbonate in T101. The regenerator strips off the absorbed CO_2. The heat for stripping is provided by the converter effluent.

The hydrogen product stream from the CO_2 absorber (T101) contains CO and CO_2 which act as hydrodesulphirizer catalyst poisons. The makeup hydrogen should typically contain less than 100 ppm CO + CO_2. In order to convert these contaminants, the hydrogen product stream is heated and passed over the methanation catalyst in R104 where the following reactions take place:

$$CO + 3H_2 \longrightarrow CH_4 + H_2O$$

$$CO_2 + 4H_2 \longrightarrow CH_4 + 2H_2O$$

5. TECHNICAL/ECONOMIC ASSESSMENT (cont'd)

5.1.2 Hydrogen Unit (cont'd)

The methanator effluent is cooled by heat exchange with its own feed in E109 and in the air cooler E110. The product containing 97 mole % H_2 is delivered to the HDS unit for make-up.

The internal heat recovery is maximized by generating steam from the furnace stack gases, the reformer effluent as well as the high temperature shift converter (R103A) effluent. The size of the equipment for a given product rate and purity varies with the carbon to hydrogen (C/H) ratio of the feed. Naphtha with a higher C/H ratio requires larger reformer furnace, shift converters and CO_2 removal system compared to the H_2 rich light gases with a low C/H ratio. In this study, a combined feedstream has been assumed. Therefore, the capital costs are higher than if all gas feed would have been assumed.

5.1.3 Gas Treating Unit

Several solvents can be used to remove H_2S from the gas streams produced in the HDS unit before being suitable as H_2 unit feedstock or as fuel gas. A lean (15-20% wt.) of monoethanolamine has been chosen on the basis of low cost and the absence of carbonyl sulphide (COS) in the gas stream.

Referring to Figure 5.1.3, the sour gases from the HDS unit are scrubbed with process water in the absorber T201 to remove ammonia. The filter coalescer aids in removing entrained hydrocarbon. The hydrogen sulphide is absorbed in T202 by a circulating solution of lean MEA. The absorber effluent gases containing only traces of H_2S are sent as feedstock to the hydrogen unit. The H_2S rich MEA from the absorber is combined with rich MEA from the recycle gas treating section of the HDS unit, heated and sent to the regenator T203. The H_2S and small amounts of hydrocarbon are stripped off from MEA with reboil heat provided in E202 by steam. The H_2S rich gas from the reflux drum (V202) is assumed routed to the Claus and Tail Gas Clean-up Units for sulphur recovery. The regenerated lean MEA solution from T203 is filtered, cooled and recycled to T202 and T2 (in the HDS unit).

The amine circulation rate is a function of the permissible pickup usually expressed in SCF of acid gas per gallon of solution. The pickup depends on the degree of regeneration and partial pressure of acid gas in the gas stream. Acid gas pickup rates of 3-4 were assumed in the study.

5.1.4 Sour Water Stripper Unit

The sour water streams produced in the HDS and gas treating units contain small quantities NH_3 and H_2S which are removed in this unit before the water can be recycled for reuse. A major portion of these acid gases are "weakly" held and can be flashed off. However, the portion that is chemically bound is stripped in the tower T-301 as shown in Figure 5.1.4.

5. TECHNICAL/ECONOMIC ASSESSMENT (cont'd)

5.1.4 Sour Water Stripper Unit (cont'd)

The amount of NH_3 and H_2S present in the stripped water are interdependent. Depending on the relative ratio, in some cases a two tower system may be employed to remove NH_3 preferentially in the first tower. The system assumed for this study would be adequate to provide less than 10 ppm H_2S and 50 mg/l. NH_3 (11) in the stripped water which can be recycled to the HDS and gas treating units. It is assumed that the small excess stream produced by dilution could be used as makeup to the deslater within the refinery.

5.2 Material Balances for the Overall Complex

Tables 5.2.2 to 5.2.17 present the material balances for the complex consisting of the residue hydrodesulphurization unit, the hydrogen unit and the gas treating and sour water stripper units. They were derived from a computer programme set up to simulate the product flash and recovery scheme shown in Figure 5.1.1. The yield information provided by the process licensors was utilized in estimating the composition of the vapour-liquid mixture at the outlet of the reactors. This formed the strarting point for the computer run. Published process information was used as far as possible supplemented by data under confidentiality agreements. The material balances derived served the following purposes:

- Establish the rates of purge gas from the recycle hydrogen circuit and the sour offgas from the product flash and fractionation sections.

- Derive the requirements of 97 mole % hydrogen required to satisfy chemical hydrogen, solution losses and purge requirements.

- Calculate the potential 97 mole % hydrogen production from the steam reforming of purge and offgases after gas treatment. In cases where these light gases could not satisfy all the feedstock requirements, naphtha was supplemented as required.

Since the cost of hydrogen is a major element in the variable cost in residue desulphurization, the accuracy of hydrogen requirements is important. The calculated hydrogen makeup requirements compare fairly well with data provided by Gulf as shown in Figure 5.2.1. However since Gulf's estimates are for end of run conditions, the size of the hydrogen unit considered in this report more closely approximate to the requirements when the catalyst has reached maximum inactivity and hydrocracking conditions begin to dominate the reactions.

The material balances also served as the basis of approximate sizing of major equipment and verifying major utility requirements. The installed cost of the facilities was estimated from the equipment list for Alberta location after due allowance for interconnecting pipeway and service facilities such as firewater and service water loops, flare blowdown facilities, control room, motor control centre, etc., which

5. TECHNICAL/ECONOMIC ASSESSMENT (cont'd)

5.2 Material Balances for the Overall Complex (cont'd)

were not included in the Process Investment Battery Limit costs provided by the licensors for U.S.A. Gulf coast location. The comparison of the independent cost estimates with those provided by the licensors enabled the identification of adjustment factors due to the inclusion of additional service facilities and the location of the complex in Edmonton compared to U.S.A. Gulf coast.

Table 5.2.18 provides a summary of material balances for the overall complex for each crude case and product sulphur level studied. The significant differences in the material balances are highlighted below:

- H-Oil does not include recycle gas treatment within the HDS unit. The acid gas removal is accomplished in the gas treating unit.

- H-Oil catalyst exhibits much higher hydrocracking compared to other processes. As shown in Table 5.2.19, the C_1-C_4 gas make is in the order of 5-7 wt % of feed for H-Oil whereas for Gulf HDS it varies from 0.56-0.82 wt % and for Chevron VRDS from 1.19 to 2.06 wt % for the Iranian Light and Canadian Light Sour Blend residue cases. As a result the Gas Treating Unit is larger in the case of H-Oil.

- Because of the large production of light gases, which preferentially constitute the hydrogen plant feedstock, no supplemental naphtha is required in almost all of the H-Oil cases. All of the fixed bed processes require the diversion of at least a portion of the naphtha product as feedstock for the hydrogen unit.

- The hydrogen makeup requirements vary considerably between processes due mainly to differences in chemical hydrogen consumption (Figures 5.2.1 and 5.2.2). The chemical hydrogen requirements for H-Oil are generally higher because of the significantly higher level of hydrocracking evident from the yield data.

- Refering to Table 5.2.19 it can be seen that there are differences in the total liquid volume yield of C_5-375°F naphtha and 375°F+ products between the processes due to variations in product densities. Gulf HDS shows the highest yields. The naphtha volume yield for H-Oil is significantly higher than for other processes thus compensating for the lower volume yield of desulphurized fuel oil.

- The sour water effluent consists of the water wash streams from the recycle gas treatment section and the product fractionator in the HDS unit. There are minor differences in the former for the different processes leading to small differences to the sour water stripper loads.

The utilities requirements for the facilities were estimated by adding the requirements for the hydrogen, gas treating and sour water stripping

5. TECHNICAL/ECONOMIC ASSESSMENT (cont'd)

5.2 Material Balances for the Overall Complex (cont'd)

units to those of the residue hydrodesulphurization unit provided by the process licensors (Appendix B). Table 5.2.20 a) to e) provide summaries of major utilities requirements for each of the crude cases per barrel of desulphurized fuel oil brokendown by individual units. As outlined in Appendix B there are some variations in the feed and product temperatures, hydrogen makeup supply pressure and steam requirements for product fractionator in the licensor supplied data. By discussions with the process licensors these differences were identified and adjustments to the overall fuel, power and steam requirements were made.

The following can be observed by a comparison of the overall utilities requirements in Tables 5.2.20 a to e.

- The power requirements are higher for H-Oil. This may be due to higher pressure operation and higher recycle gas circulation rates.

- The fuel requirements in the HDS unit are higher for H-Oil for Iranian Light cases. This is likely due to the inclusion of a vacuum tower for an internal gas oil recycle stream generation. The marginally higher hydrogen plant capacities also contribute to increased utility consumption in the H-Oil cases.

- There seem to be minor variations amongst the other processes though it should be emphasized that the data provided by the licensors have not been derived from detailed process optimization studies.

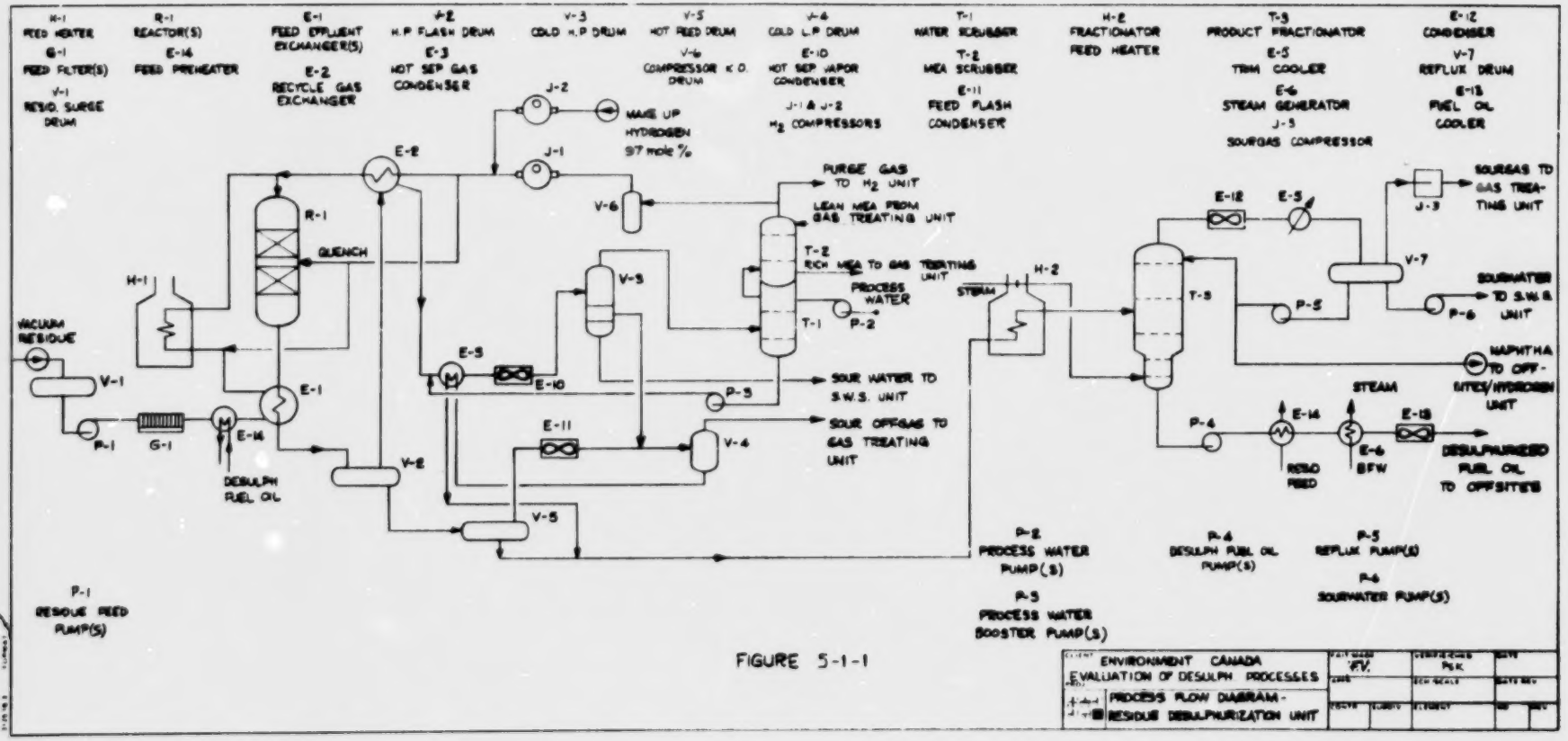


FIGURE 5-1-1

CLIENT	ENVIRONMENT CANADA
PROJECT	EVALUATION OF DESULPH PROCESSES
TITLE	PROCESS FLOW DIAGRAM - RESIDUE DESULPHURIZATION UNIT
DRAWN	FV.
CHECKED	PSK
SCALE	
DATE	

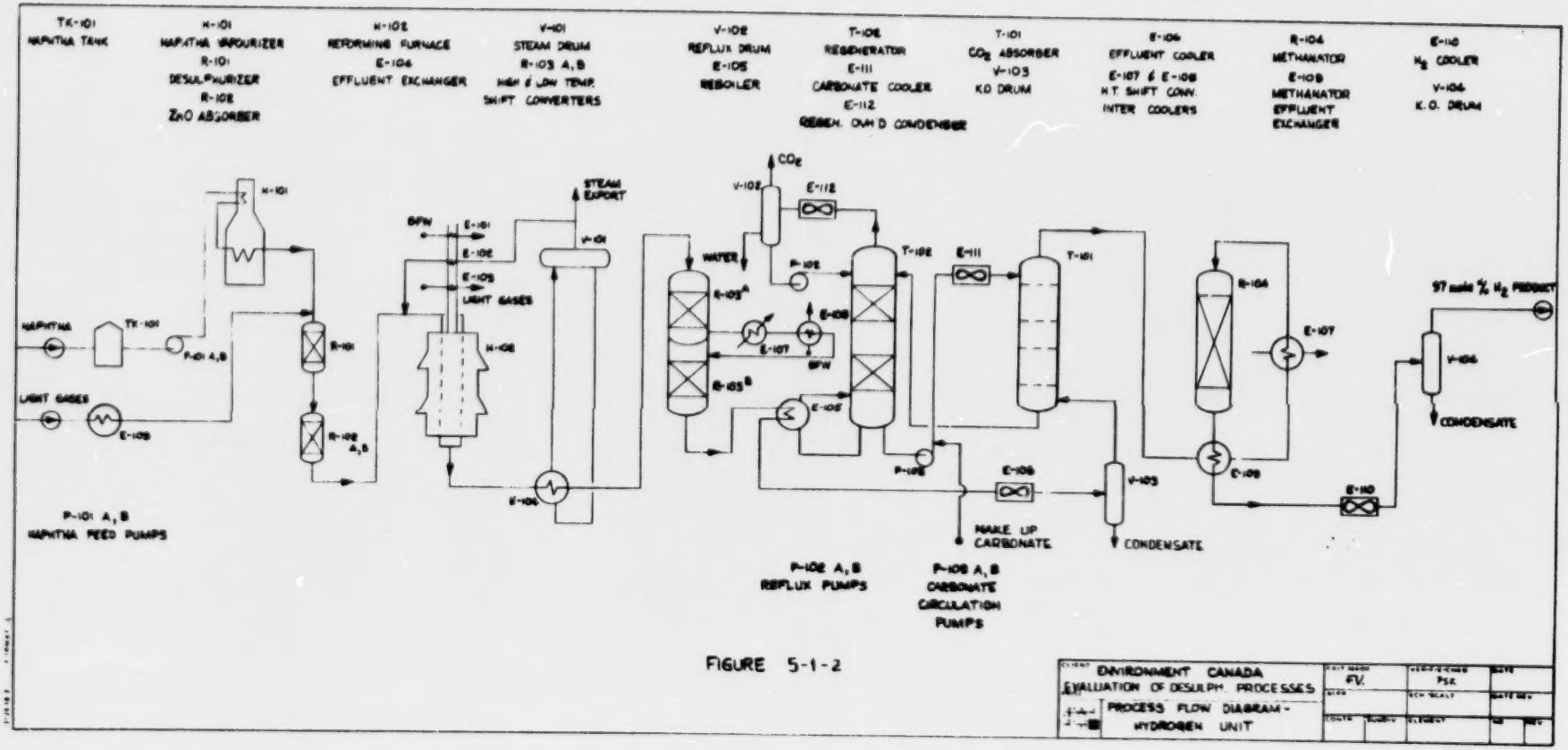


FIGURE 5-1-2

ENVIRONMENT CANADA	FV.	PSK.	
EVALUATION OF DESULPH. PROCESSES			
PROCESS FLOW DIAGRAM - HYDROGEN UNIT			

T-201
WATER SCRUBBER
G-201
FILTER COALESCER

E-203
LEAN MEA
COOLER

T-202
MEA ABSORBER
V-201
MEA SURGE DRUM
G-202
LEAN MEA FILTERS

T-203
MEA REGENERATOR
V-202
REFLUX DRUM

E-201
CONDENSER
E-202
REBOILER

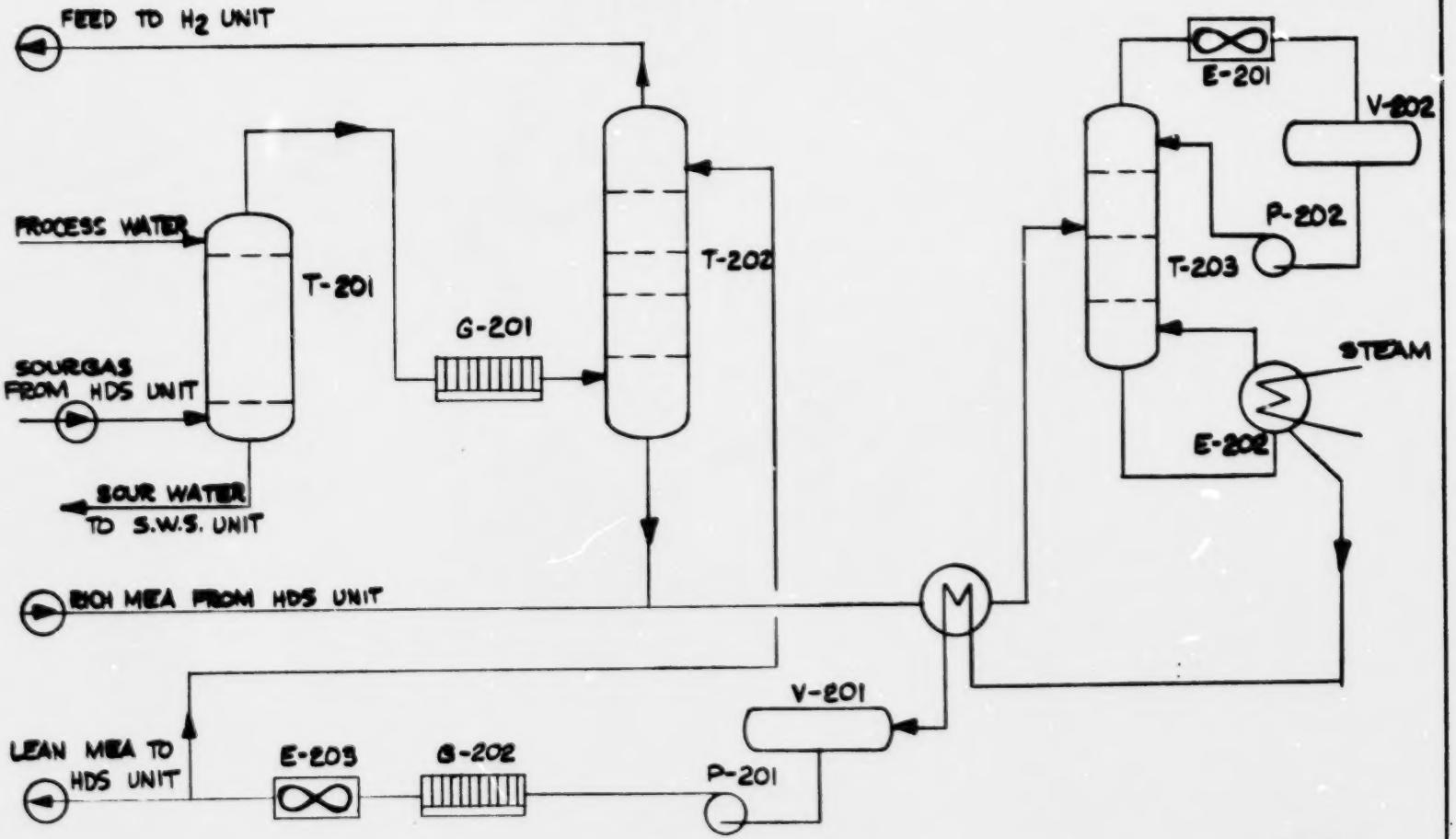


FIGURE 5-1-3

P-201
LEAN MEA PUMP(S)

P-202
REFLUX PUMP(S)

CLIENT ENVIRONMENT CANADA
PROJ. EVALUATION OF DESULPH. PROCESSES

PROCESS FLOW DIAGRAM - GAS TREATING UNIT

FAIT MADE F.V.
APPR.
CONTR.
SUBDIV.
VERIFIE/CHKD PSK
ECH./SCALE
ELEMENT
NO.
DATE REV.
DATE
REV.

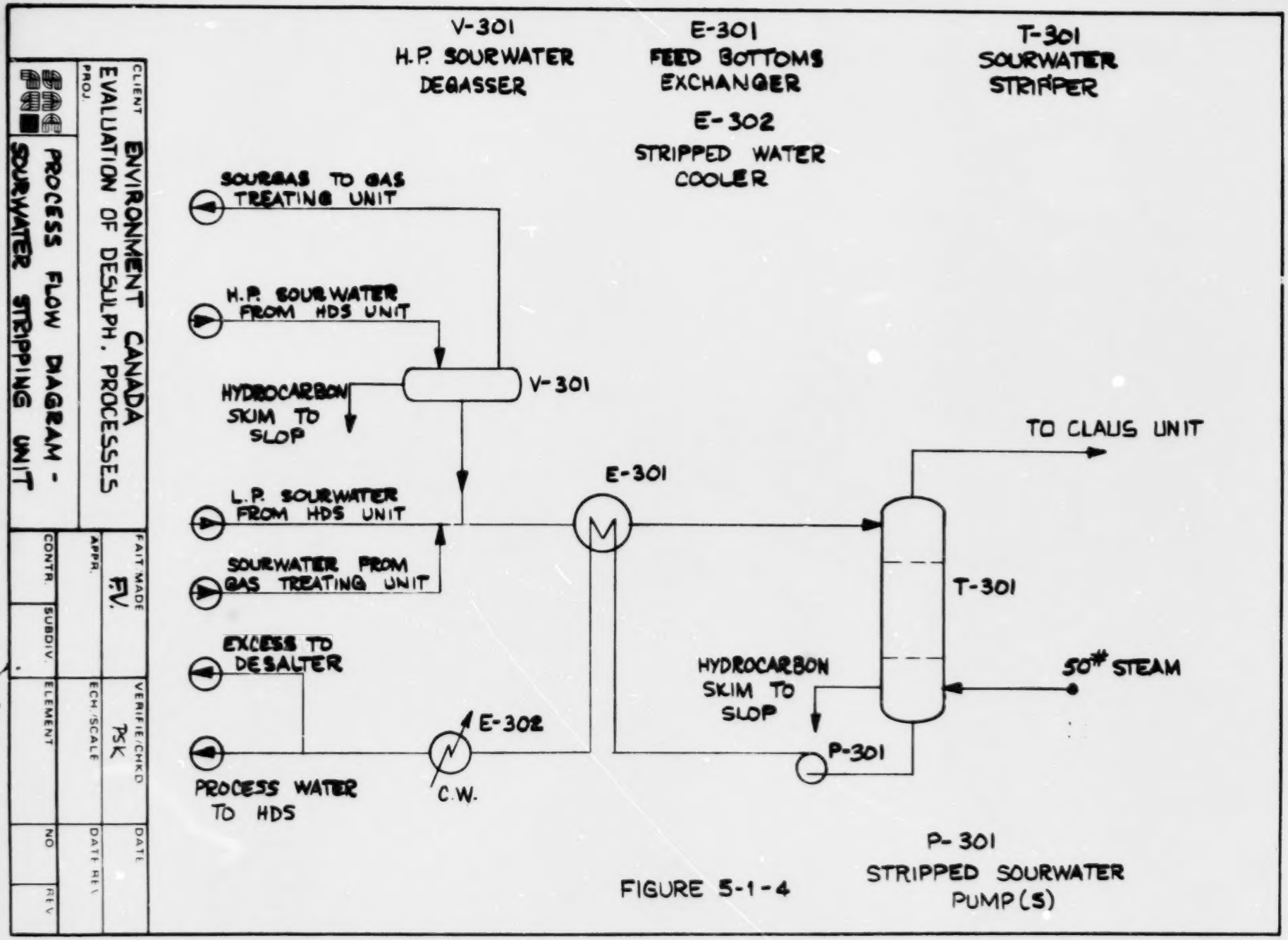

V-301
H.P. SOURWATER
DEGASSER
E-301
FEED BOTTOMS
EXCHANGER
E-302
STRIPPED WATER
COOLER
T-301
SOURWATER
STRIPPER
SOURGAS TO GAS
TREATING UNIT
H.P. SOURWATER
FROM HDS UNIT
HYDROCARBON
SKIM TO
SLOP
V-301
L.P. SOURWATER
FROM HDS UNIT
SOURWATER FROM
GAS TREATING UNIT
E-301
TO CLAUS UNIT
T-301
EXCESS TO
DESALTER
HYDROCARBON
SKIM TO
SLOP
50# STEAM
E-302
P-301
PROCESS WATER
TO HDS
C.W.
P-301
STRIPPED SOURWATER
PUMP(S)
CLIENT
ENVIRONMENT CANADA
EVALUATION OF DESULPH. PROCESSES
PROJ.
PROCESS FLOW DIAGRAM -
SOURWATER STRIPPING UNIT
CONTR.
APPR.
FAIT MADE
F.V.
SUBDIV.
ELEMENT
ECH/SCALE
VERIFIE/CHK'D
PSK
NO
DATE REV
DATE
REV

FIGURE 5-1-4

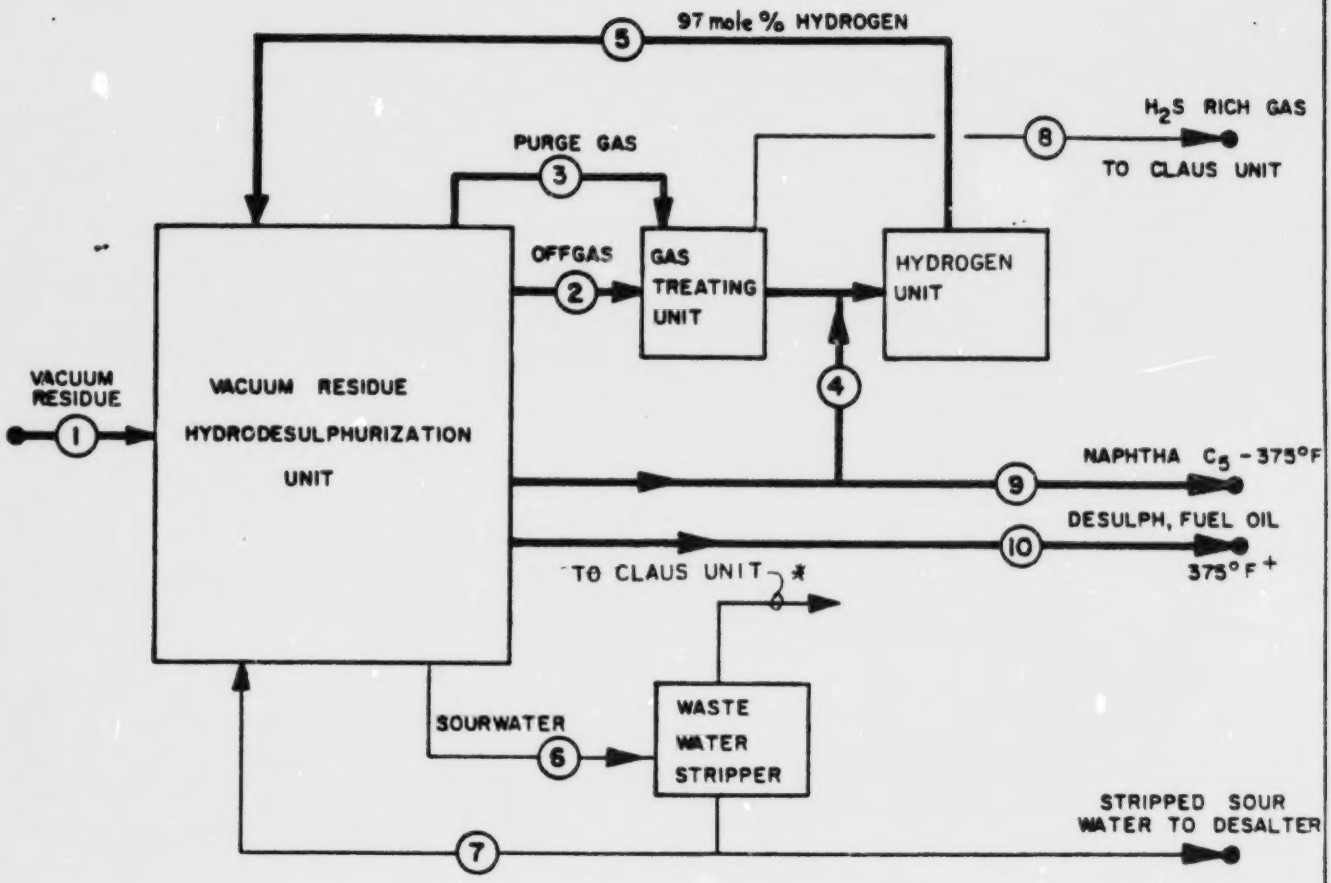
TABLE 5.2.1

Comparison of Hydrogen Requirements
Calculated vs. Makeup Licensor (Gulf) Data

(Basis: 20,000 BPSD)

	Wt.%S in 375°F* Product	97 mole % H_2 Make Up (MMSCFD) Calculated	Gulf *
Iranian Light	1.0	14.66	15.21
Iranian Light	0.5	15.00	18.11
Canadian Light Sour Blend	1.0	12.60	12.96
	0.5	17.17	16.17

* End of run conditions and include 60 SCFB for mechanical losses. Gulf's figures adjusted for 97 mole % purity.



* (856 lb/hr.)

STREAM*	1	2 - 3	4	5	6	7	8	9	10	
IDENTIFICATION	Residue	Purge & Offgas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Water Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil	
LIQUID FLOW BPSD	20,000	-	241	-	2,014	1,371	-	1,252	19,280	
GAS FLOW MMSCFD	-	10.94		22.90	-	-	1.73	-	-	
lb/hr	298,800	16,334	2,700	6,432	28,192	20,000	6,614	14,032	271,310	

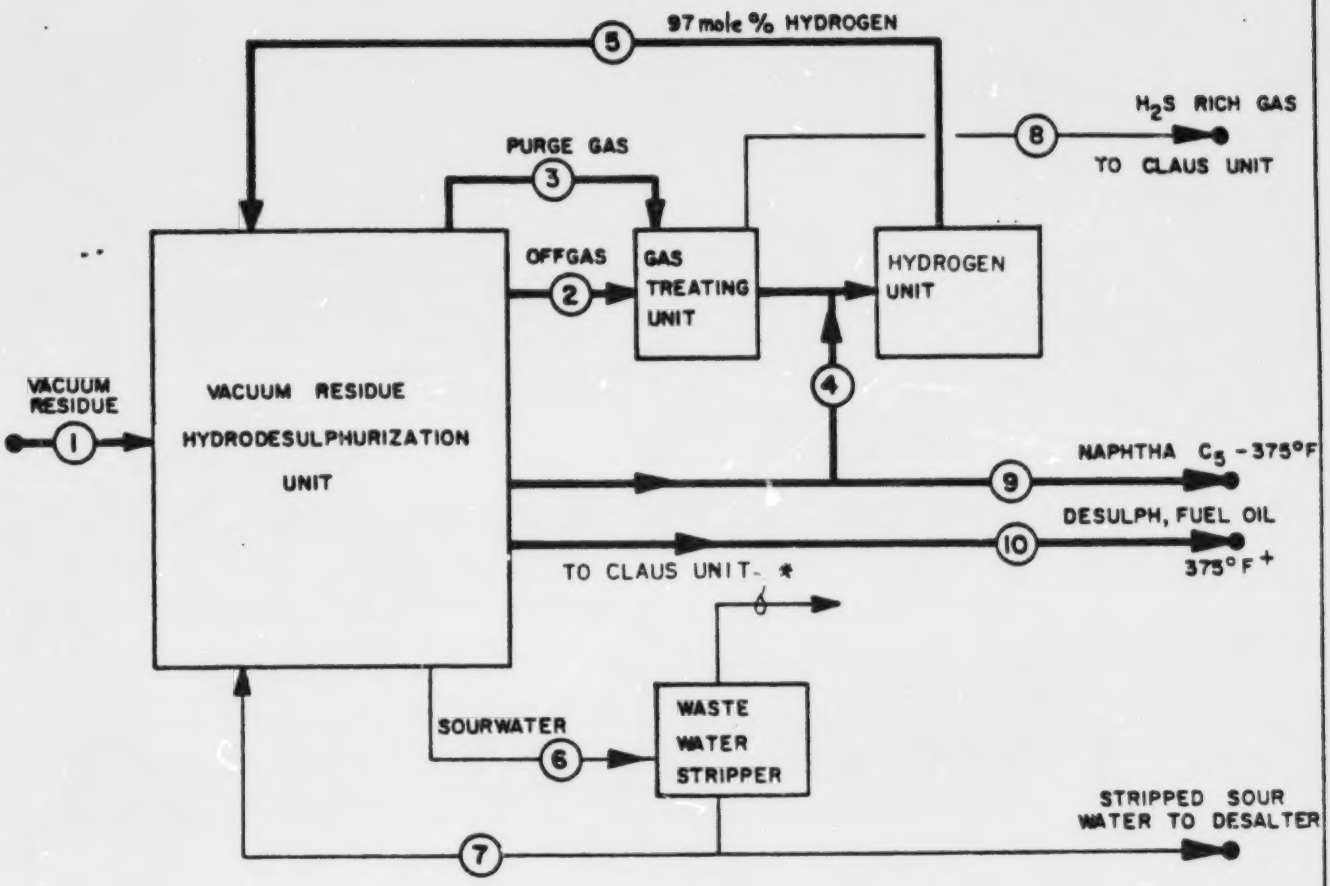
CRUDE SOURCE: Tia Juana

PROCESS: H-Oil

WT %S IN 375$^+$ PRODUCT: 1.0

TABLE 5.2.2

CLIENT ENVIRONMENT CANADA		FAIT/MADE F.V.	VERIFIE/CHKD	DATE	
PROJ. EVALUATION OF DESULPHURIZATION PROCESSES		APPR.	ECH./SCALE	DATE REV.	
MATERIAL BALANCE FOR THE OVERALL COMPLEX		CONTR. / SUBDIV.	ELEMENT	NO.	REV.



* (1062 lb/hr.)

STREAM*	1	2 3	4	5	6	7	8	9	10	
IDENTIFICATION	Residue	Purge & Offgas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Water Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil	
LIQUID FLOW BPSD	20,000	-	-	-	2,028	1,371	-	2,768	18,420	
GAS FLOW MMSCFD	-	12.69	-	32.00	-	-	2.14	-	-	
lb/hr	298,800	22,864	-	8,990	28,398	20,000	8,201	28,385	253,680	

CRUDE SOURCE: Tia Juana

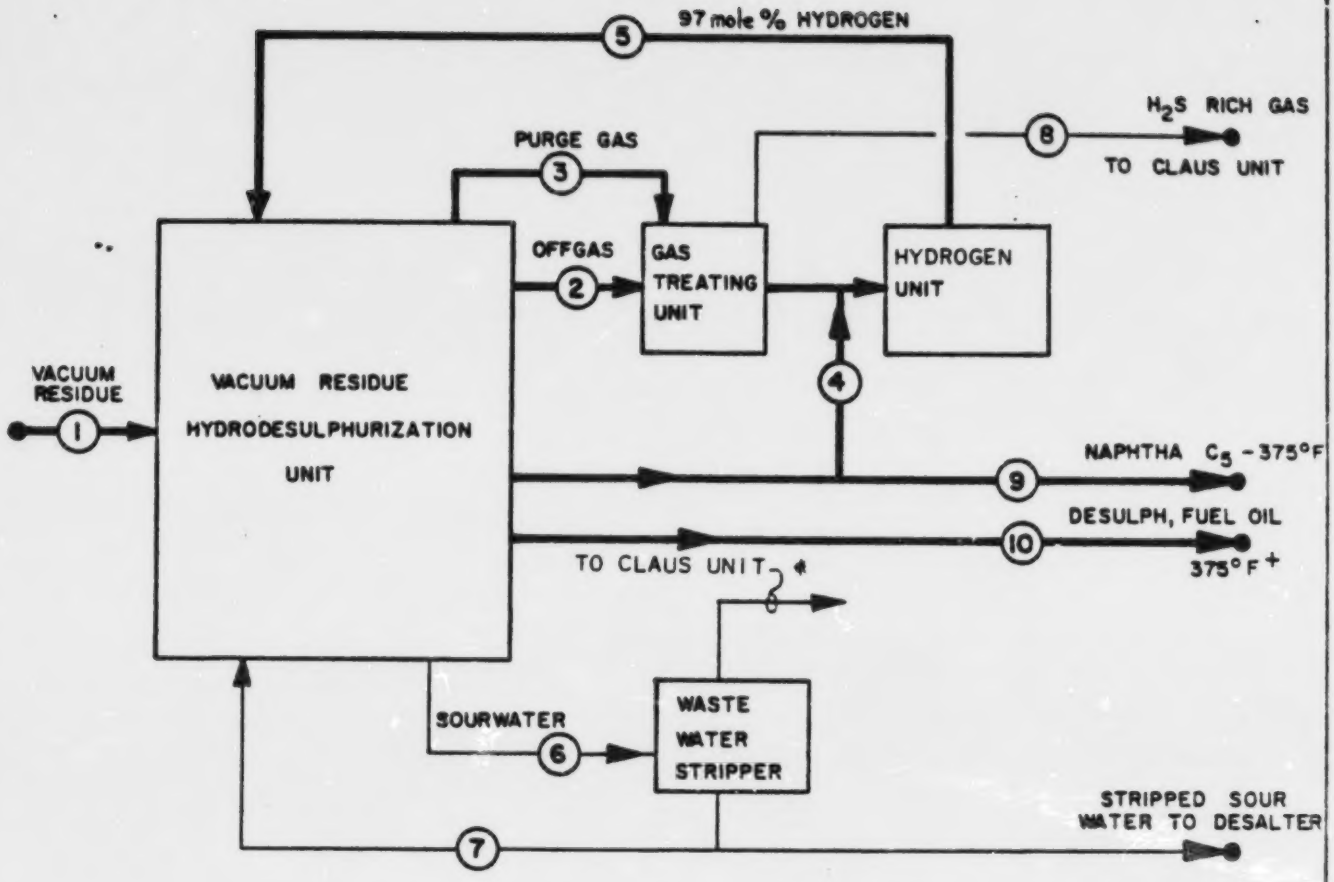
PROCESS: H-Oil

WT % S IN 375$^+$ PRODUCT: 0.5

TABLE 5.2.3

CLIENT ENVIRONMENT CANADA	FAIT/MADE F.V.	VERIFIE/CHKD	DATE
PROJ. EVALUATION OF DESULPHURIZATION PROCESSES	APPR.	ECH./SCALE	DATE REV
MATERIAL BALANCE FOR THE OVERALL COMPLEX	CONTR. / SUBDIV.	ELEMENT	NO. / REV

3126-02/2 FORMAT



* (794 lb/hr.)

STREAM*	1	2 - 3	4	5	6	7	8	9	10	
IDENTIFICATION	Residue	Purge & Offgas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Water	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil	
LIQUID FLOW BPSD	20,000	-	-	-	2,000	1,372	-	1,280	19,320	
GAS FLOW MMSCFD	-	9.10	-	20.00	-	-	1.55	-	-	
lb/hr	297,555	15,148	-	5,620	28,017	20,000	6,050	13,480	273,750	

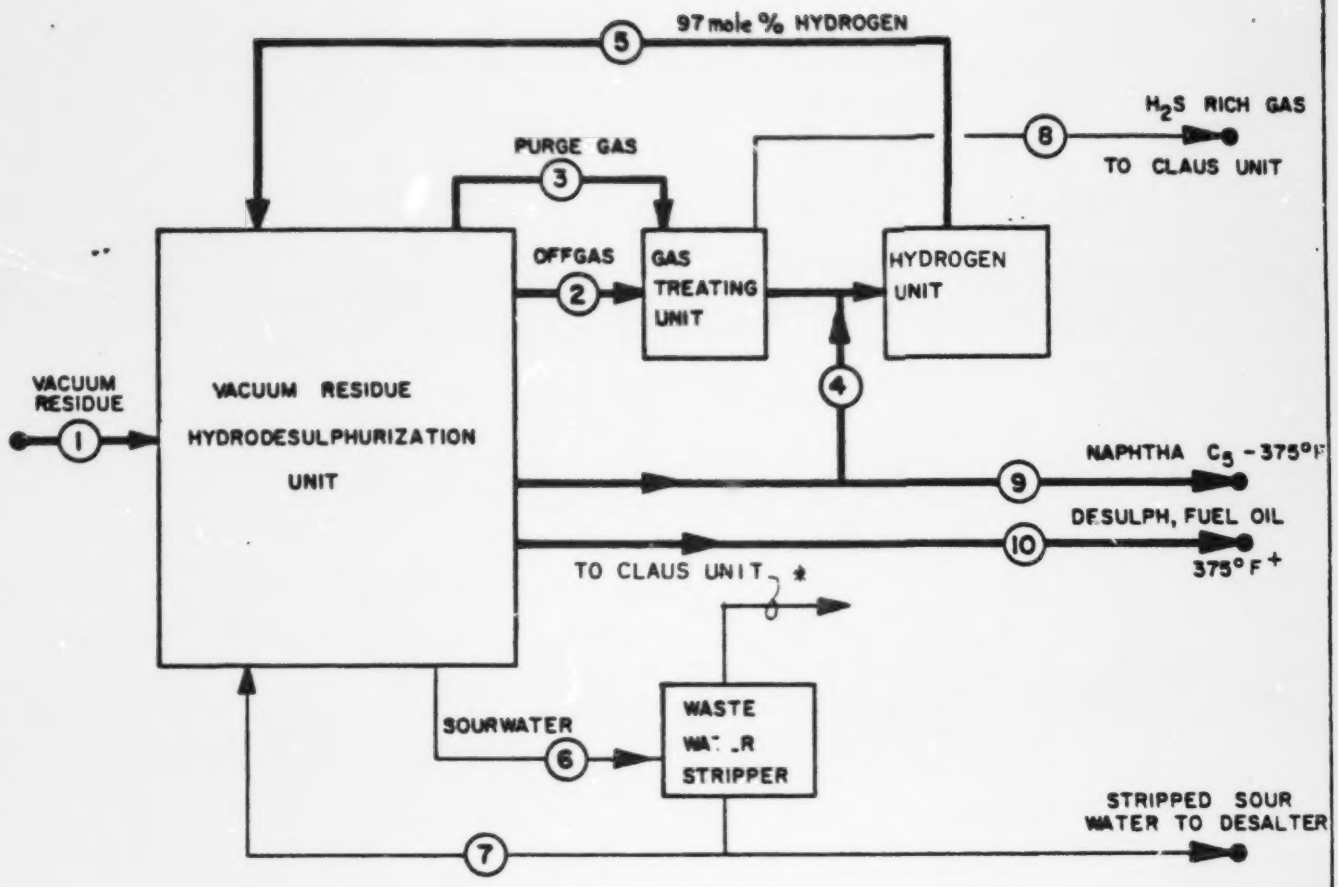
CRUDE SOURCE: Iranian Light

PROCESS: H-Oil

WT %S IN 375+ PRODUCT: 1.0

TABLE 5.2.4

CLIENT ENVIRONMENT CANADA		FAIT/MADE F.V.	VERIFIE/CHKD	DATE	
PROJ. EVALUATION OF DESULPHURIZATION PROCESSES		APPR.	ECH./SCALE	DATE REV.	
MATERIAL BALANCE FOR THE OVERALL COMPLEX		CONTR. / SUBDIV.	ELEMENT	NO.	REV.



STREAM *	1	2 - 3	4	5	6	7	8	9	10	
IDENTIFICATION	Residue	Purge & Offgas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Water	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil	
LIQUID FLOW BPSD	20,000	-	-	-		1,372	-	1,718	19,060	
GAS FLOW MMSCFD	-	11.39	-	25.00	-	-	2.07	-	-	
lb/hr	297,555	19,399	-	7,020	28,360	20,000	7,913	18,148	266,004	

CRUDE SOURCE: Iranian Light

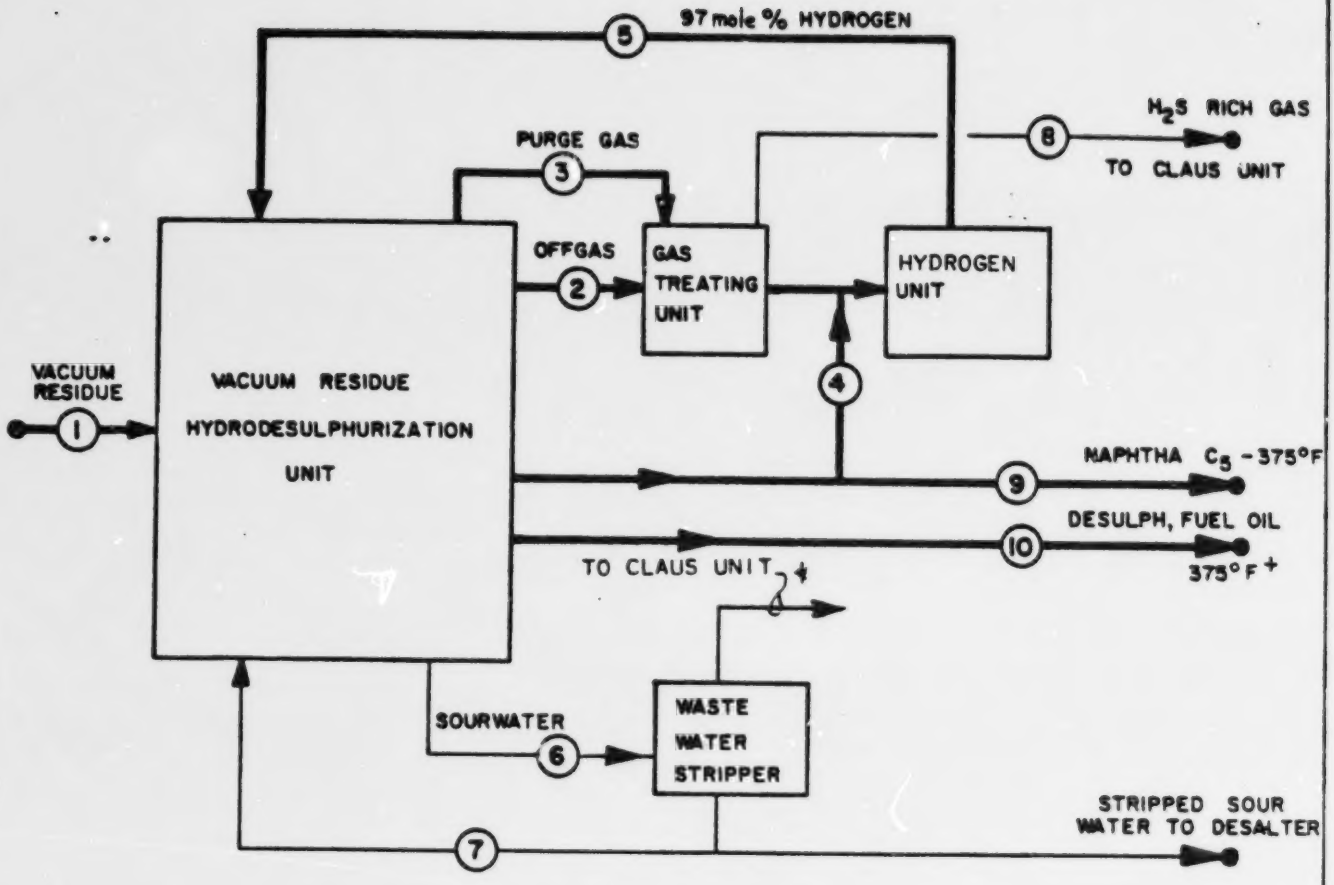
PROCESS: H-Oil

WT %S IN 375$^+$ PRODUCT: 0.5

TABLE 5.2.5

CLIENT ENVIRONMENT CANADA	FAIT/MADE F.V.	VERIFIE/CHKD	DATE	
PROJ EVALUATION OF DESULPHURIZATION PROCESSES	APPR.	ECH./SCALE	DATE REV.	
MATERIAL BALANCE FOR THE OVERALL COMPLEX	CONTR. / SUBDIV.	ELEMENT	NO.	REV.

FORMAT 315 U2/2



* (1228 lb/hr.)

STREAM*	1	2 - 3	4	5	6	7	8	9	10	
IDENTIFICATION	Residue	Purge & Offgas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Sour Water	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil	
LIQUID FLOW BPSD	20,000	-	-	-	1,754	1,097	-	1,411	19,300	
GAS FLOW MMSCFD	-	9.76	-	20.0	-	-	1.93	-	-	
lb/hr	297,500	16,889	-	5,620	24,564	16,000	7,400	15,170	269,832	

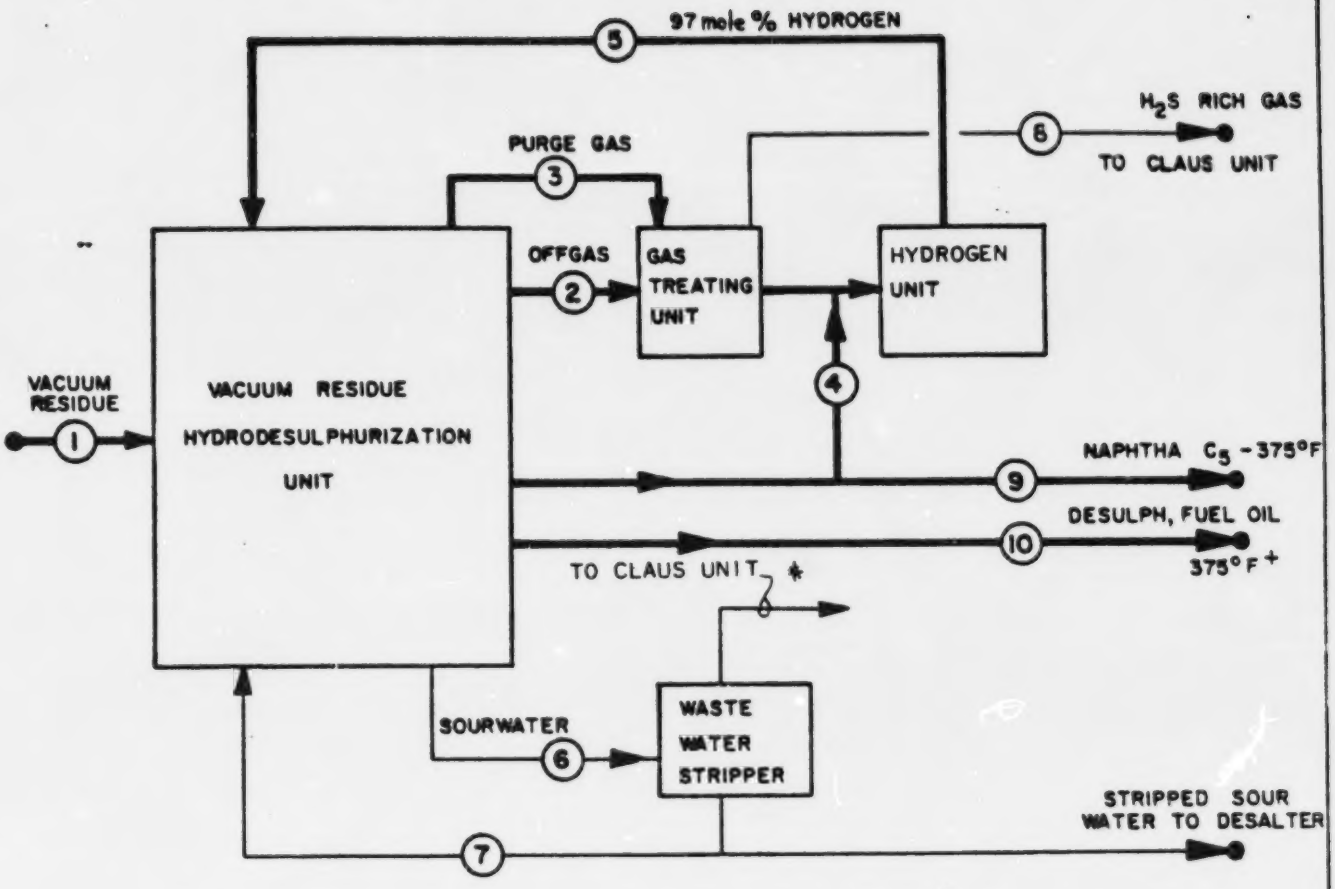
CRUDE SOURCE: Canadian Light Sour Blend

PROCESS: H-Oil

WT %S IN 375$^+$ PRODUCT: 1.0

TABLE 5.2.6

CLIENT ENVIRONMENT CANADA		FAIT/MADE F.V.		VERIFIE/CHKD	DATE
PROJ EVALUATION OF DESULPHURIZATION PROCESSES		APPR.		ECH./SCALE	DATE REV.
MATERIAL BALANCE FOR THE OVERALL COMPLEX		CONTR.	SUBDIV.	ELEMENT	NO. / REV



* (838 lb/hr.)

STREAM**	1	2 - 3	4	5	6	7	8	9	10	
IDENTIFICATION	Residue	Purge & Offgas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Water	Class Unit Feed	Excess Naphtha	Desulph. Fuel Oil	
LIQUID FLOW BPSD	20,000	-	-	-	1,707	1,097	-	2,352	18,700	
GAS FLOW MMSCFD	-	11.80	-	26.80	-	-	2.34	-	-	
lb/hr	297,500	23,328	-	7,527	23,900	16,000	8,956	25,585	255,850	

CRUDE SOURCE: Canadian Light Sour Blend

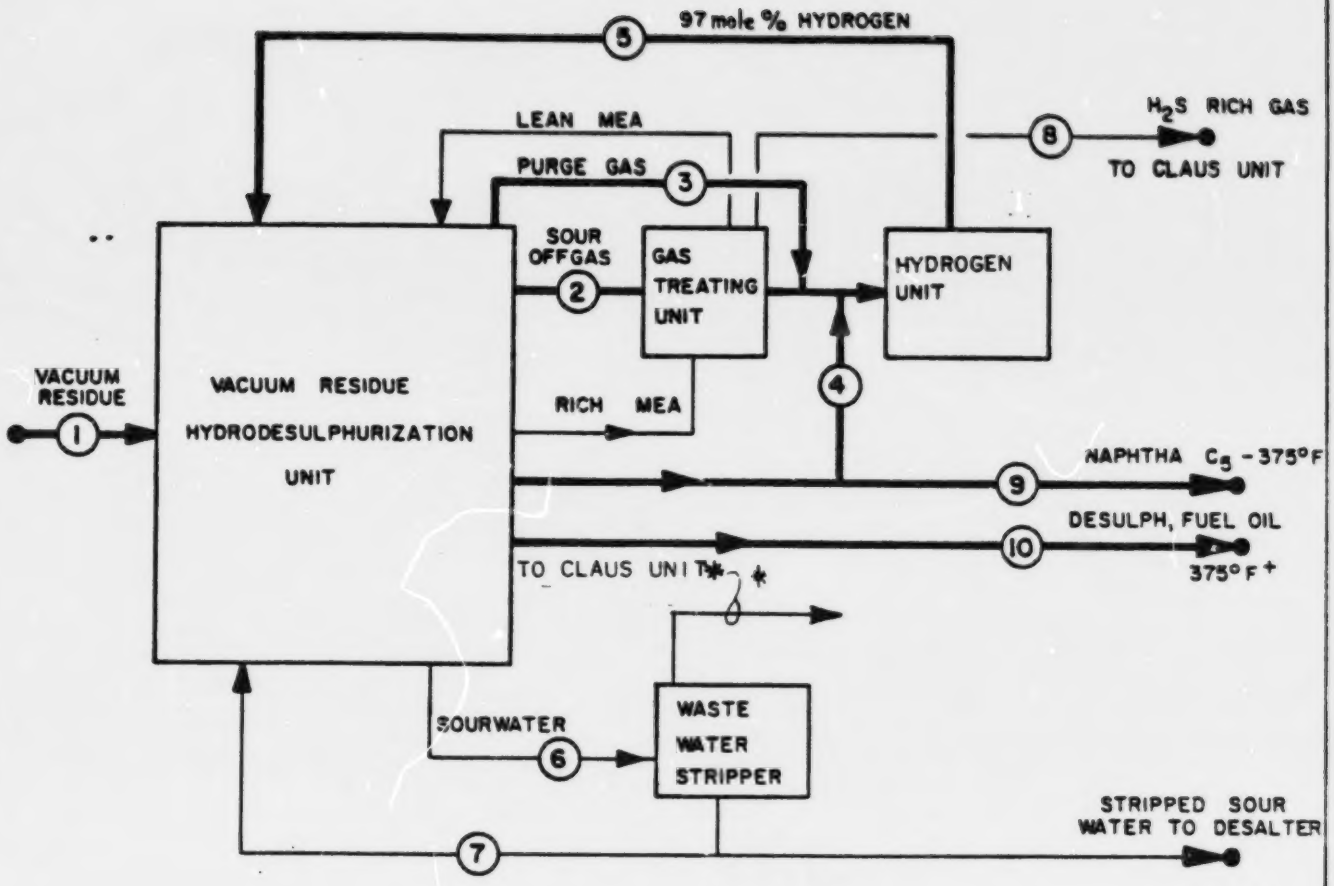
PROCESS: H-Oil

WT %S IN 375+ PRODUCT: 0.5

TABLE 5.2.7

CLIENT ENVIRONMENT CANADA		FAIT/MADE F.V.	VERIFIE/CHKD	DATE
PROJ EVALUATION OF DESULPHURIZATION PROCESSES		APPR.	ECH./SCALE	DATE REV.
MATERIAL BALANCE FOR THE OVERALL COMPLEX		CONTR. / SUBDIV.	ELEMENT	NO. / REV

FORMAT :
3125 02/2



* (904 lb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped S.W. Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	143	-	1,406	800	-	213	19,980
GAS FLOW MM3CFD	-	2.77	4.87	-	19.20	-	-	1.86	-	-
lb/hr	297,555	3,638	3,697	1,640	5,393	19,686	11,667	,111	2,447	284,724

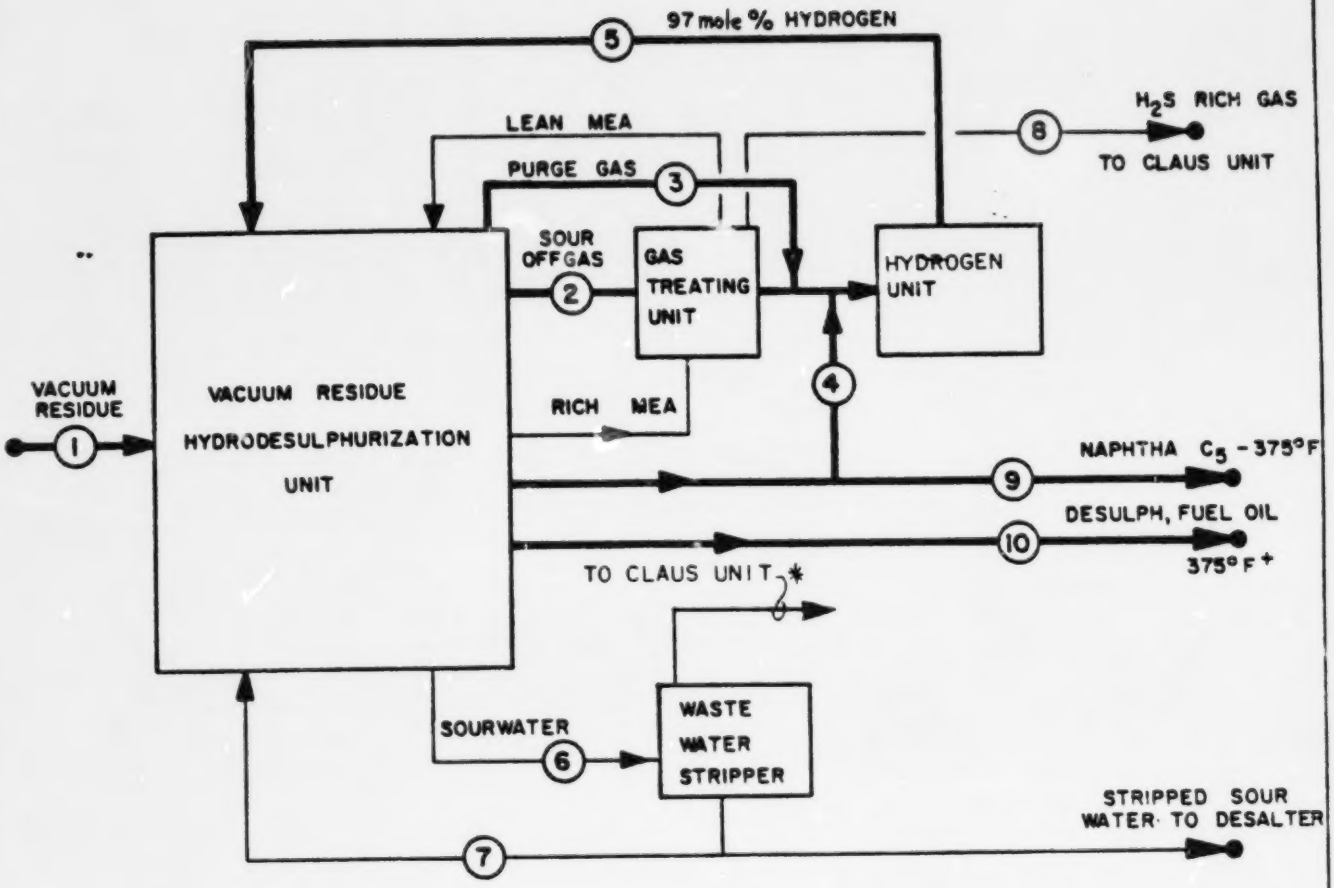
CRUDE SOURCE: Iranian Light

PROCESS: Chevron VRDS

WT %S IN 375+ PRODUCT: 1.0

TABLE 5.2.8

CLIENT ENVIRONMENT CANADA	FAIT/MADE F.V.	VERIFIE/CHKD	DATE
PROJ EVALUATION OF DESULPHURIZATION PROCESSES	APPR.	ECH. SCALE	DATE REV
MATERIAL BALANCE FOR THE OVERALL COMPLEX	CONTR. / SUBDIV.	ELEMENT	NO. / REV



* (1036 lb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped S.W. Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	462	-	2,273	1,624	-	44	19,880
GAS FLOW MMSCFD	-	2.83	5.71	-	28.00	-	-	2.25	-	-
lb/hr	297,555	3,893	5,538	5,260	7,864	31,825	23,680	8,634	500	281,718

CRUDE SOURCE: Iranian Light

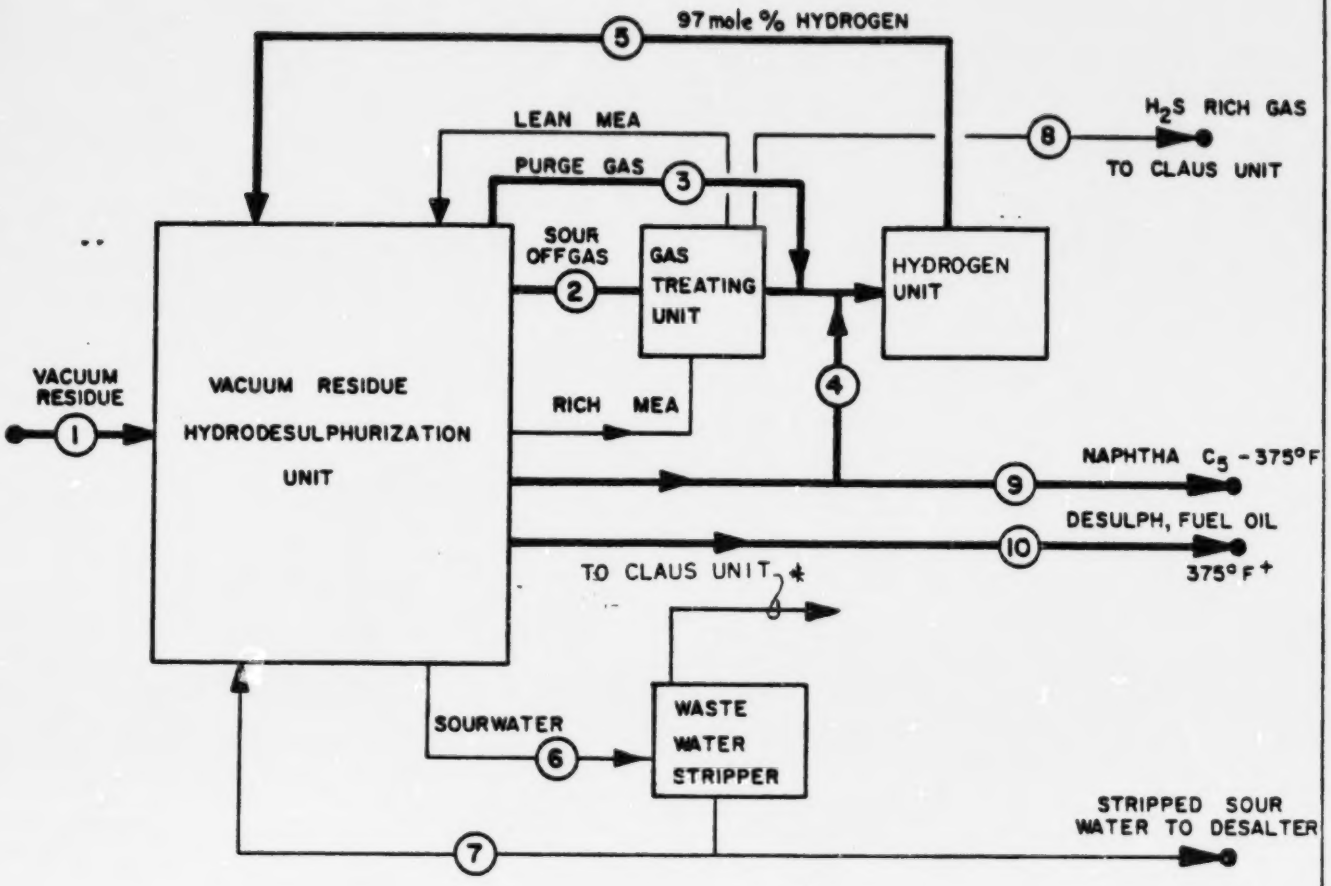
PROCESS: Chevron VRDS

WT %S IN 375+ PRODUCT: 0.5

TABLE 5.2.9

CLIENT ENVIRONMENT CANADA	FAIT/MADE F.V.	VERIFIE/CHKD	DATE
PROJ. EVALUATION OF DESULPHURIZATION PROCESSES	APPR.	ECH./SCALE	DATE REV.
MATERIAL BALANCE FOR THE OVERALL COMPLEX	CONTR. / SUBDIV.	ELEMENT	NO. / REV.

FORMAT 3126.02/2



* (1263 lb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped S.W. Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	137	-	2,164	1,487	-	332	19,920
GAS FLOW MMSCFD	-	2.49	5.63	-	21.5	-	-	1.92	-	-
lb/hr	297,500	3,492	4,953	1,545	6,040	30,301	21,680	7,366	3,744	282,167

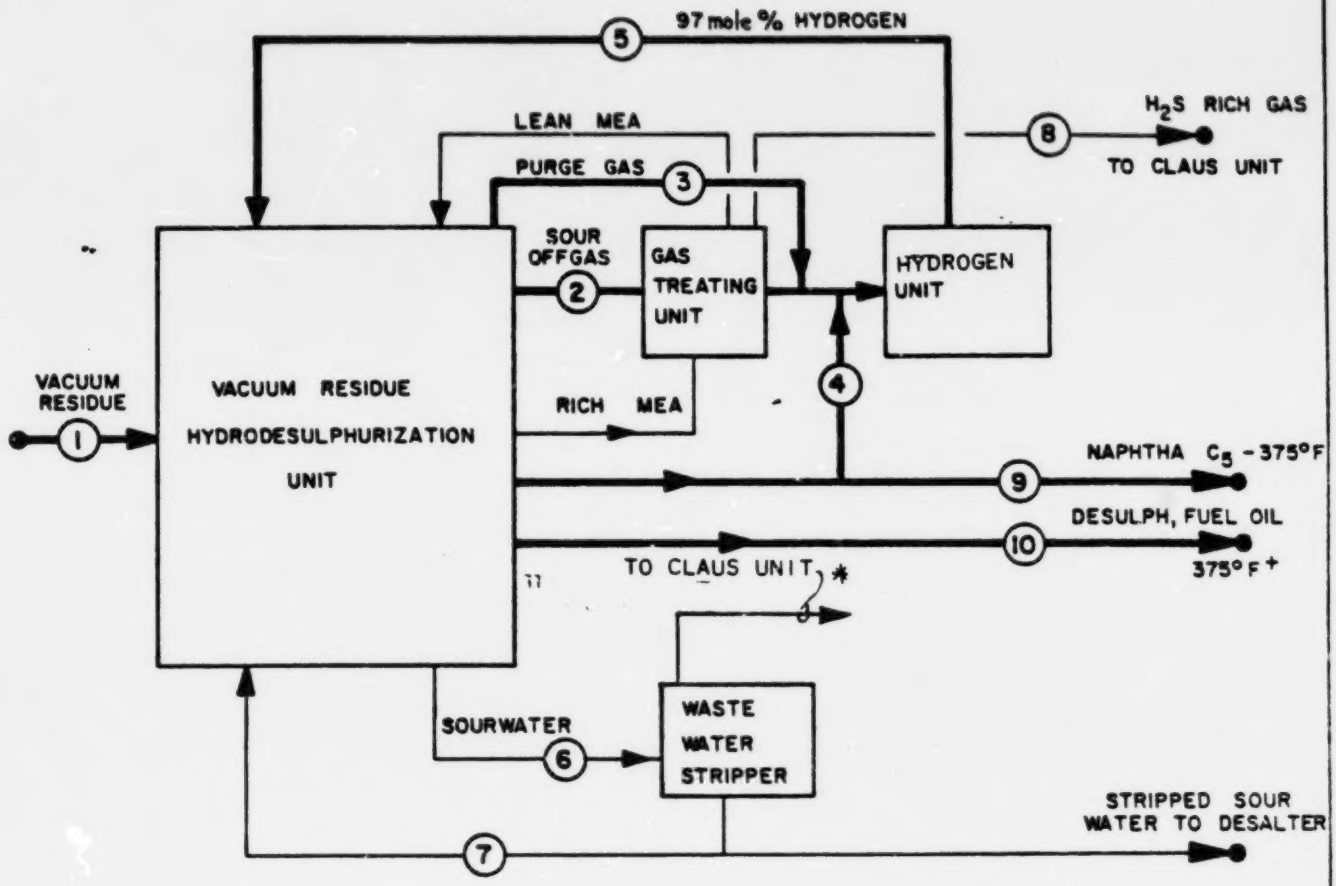
CRUDE SOURCE: Canadian Light Sour Blend

PROCESS: Chevron VRDS

WT %S IN 375+ PRODUCT: 1.0

TABLE 5.2.10

CLIENT ENVIRONMENT CANADA		FAIT/MADE F.V.	VERIFIE/CHKD	DATE	
PROJ. EVALUATION OF DESULPHURIZATION PROCESSES		APPR.	ECH. SCALE	DATE REV.	
MATERIAL BALANCE FOR THE OVERALL COMPLEX		CONTR. / SUBDIV.	ELEMENT	NO.	REV



* (2094 lb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped S.W. Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	306	-	3,773	2,973	-	366	19,780
GAS FLOW MMSCFD	-	2.61	6.58	-	31.00	-	-	2.23	-	-
lb/hr	297,500	3,967	6,910	3,460	8,710	52,817	43,355	8,557	4,137	278,065

CRUDE SOURCE: Canadian Light Sour Blend

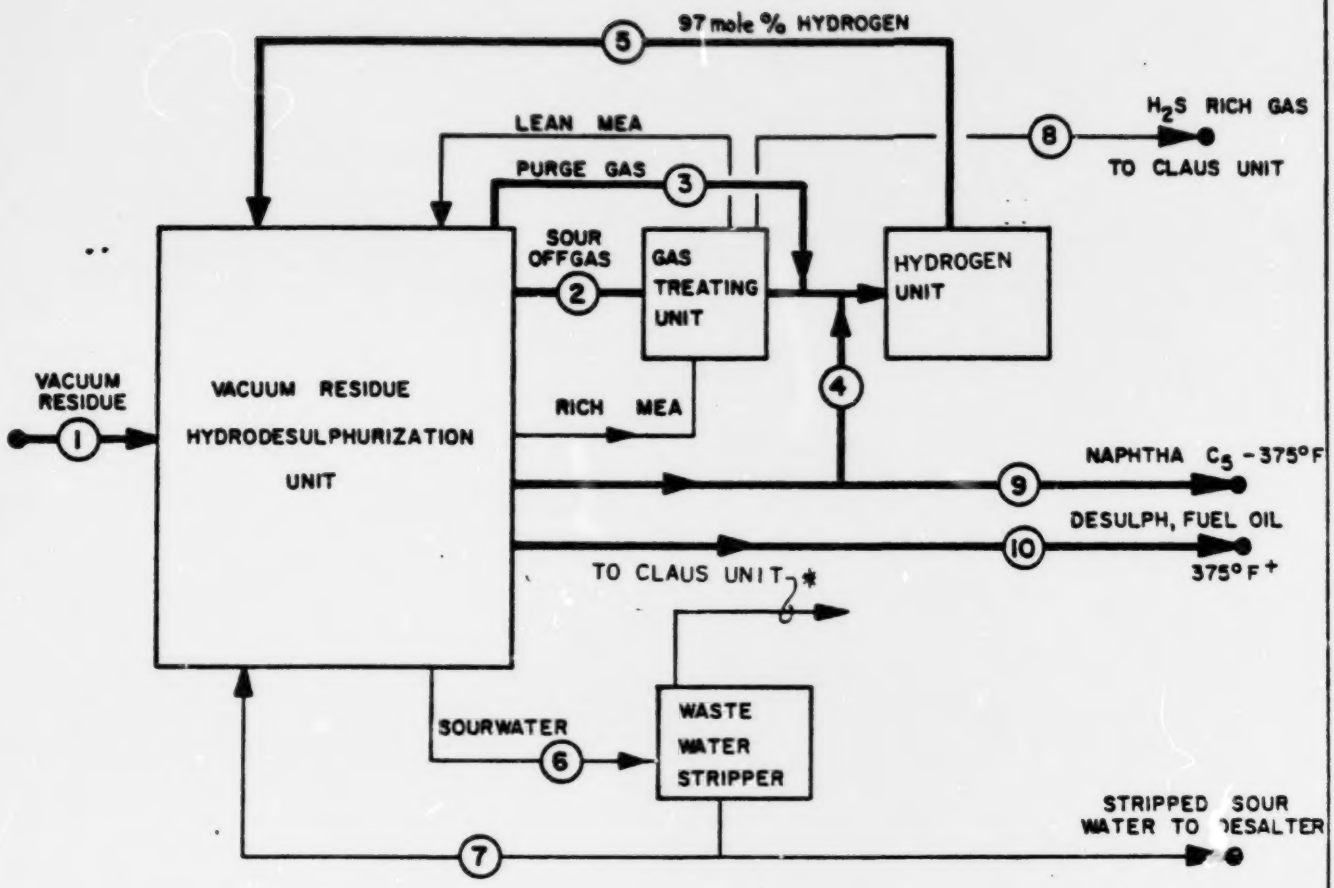
PROCESS: Chevron VRDS

WT %S IN 375+ PRODUCT: 0.5

TABLE 5.2.11

CLIENT ENVIRONMENT CANADA		FAIT/MADE F.V.	VERIFIE/CHKD	DATE	
PROJ. EVALUATION OF DESULPHURIZATION PROCESSES		APPR.	ECH./SCALE	DATE REV.	
MATERIAL BALANCE FOR THE OVERALL COMPLEX		CONTR.	SUBDIV. ELEMENT	NO.	REV

FORMAT 3125.02/2



* (1147 lb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped S.W. Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	279	-	2,936	2,743	-	142	20.950
GAS FLOW MMSCFD	-	4.04	2.97	-	14.66	-	-	1.52	-	-
lb/hr	297,555	3,523	1,742	3,085	4,118	41,098	40,000	5,816	1,568	285,876

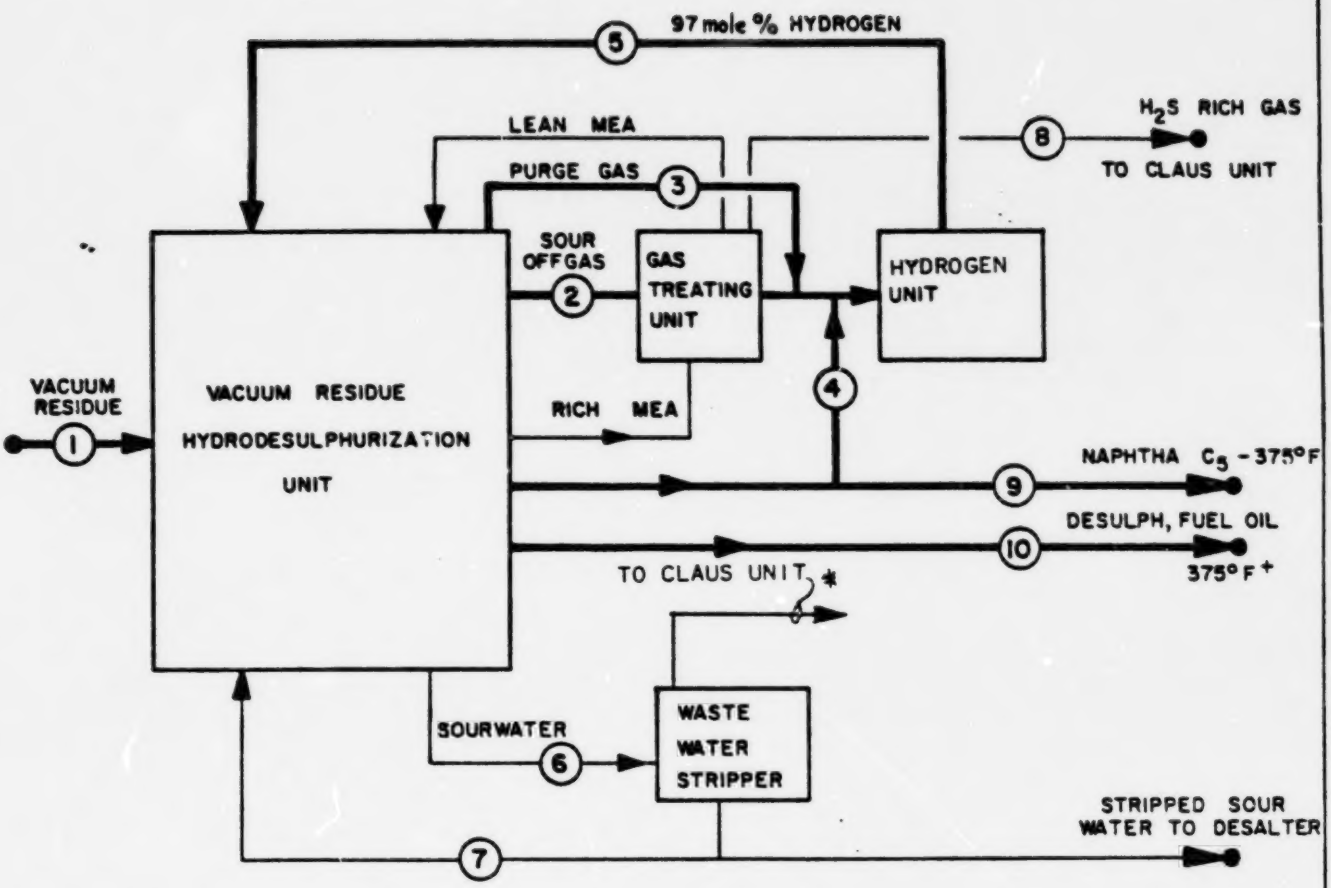
CRUDE SOURCE: Iranian Light

PROCESS: Gulf HDS II

WT %S IN 375+ PRODUCT: 1.0

TABLE 5.2.12

CLIENT ENVIRONMENT CANADA		FAIT/MADE F.V.		VERIFIE/CHKD	DATE
PROJ. EVALUATION OF DESULPHURIZATION PROCESSES		APPR.		ECH./SCALE	DATE REV.
MATERIAL BALANCE FOR THE OVERALL COMPLEX		CONTR.	SUBDIV.	ELEMENT	NO. / REV.



* (1328 lb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Water	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	362	-	3,885	3,634	-	222	20,378
GAS FLOW MMSCFD	-	2.97	4.47	-	19.08	-	-	1.92	-	-
lb/hr	297,555	3,777	2,618	4,055	5,360	54,383	53,000	7,359	2,487	282,313

CRUDE SOURCE: Iranian Light

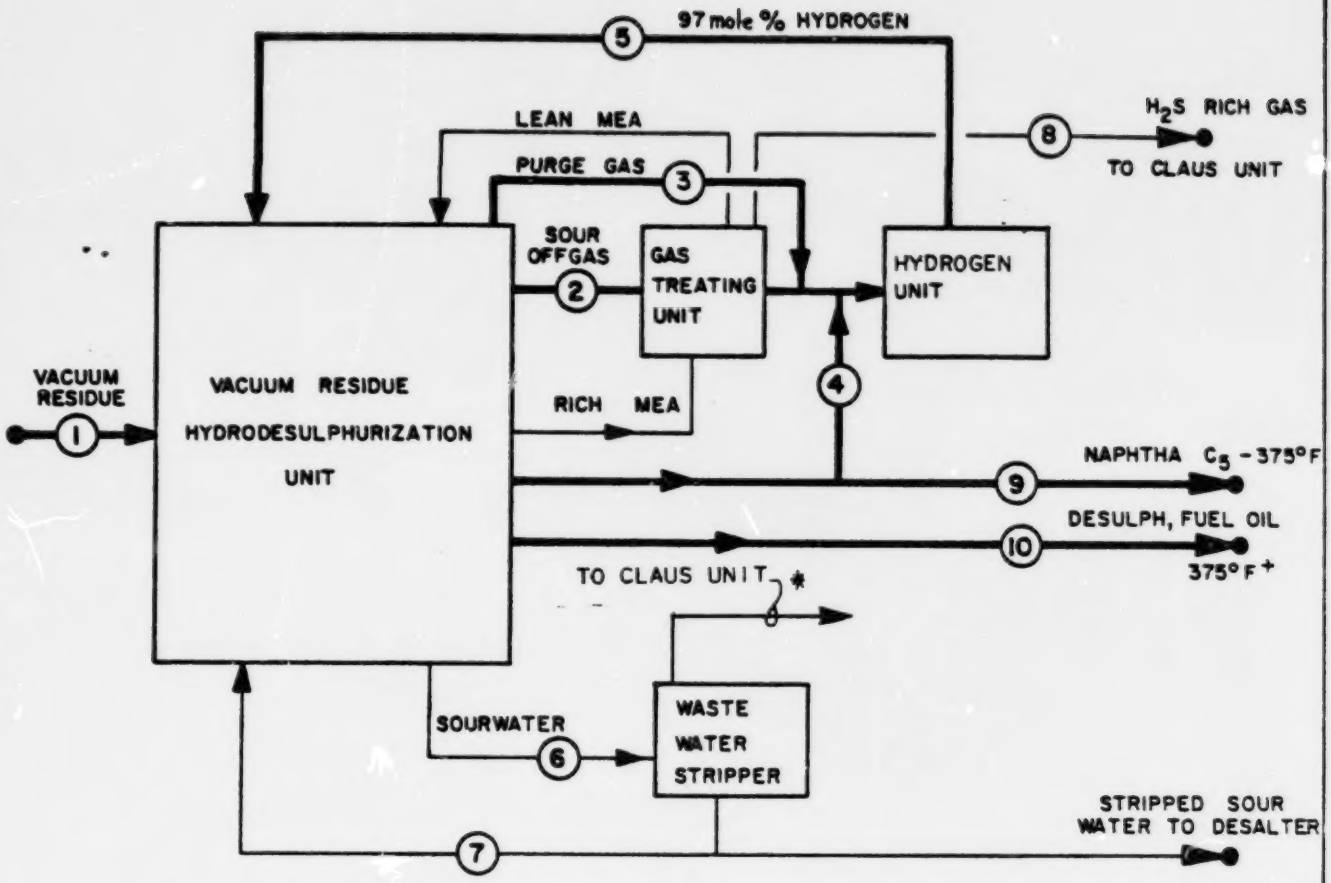
PROCESS: Gulf HDS II

WT %S IN 375+ PRODUCT: 0.5

TABLE 5.2.13

CLIENT: ENVIRONMENT CANADA	FAIT/MADE: F.V.	VERIFIE/CHKD	DATE
PROJ: EVALUATION OF DESULPHURIZATION PROCESSES	APPR.	ECH./SCALE	DATE REV.
MATERIAL BALANCE FOR THE OVERALL COMPLEX	CONTR. / SUBDIV.	ELEMENT	NO. / REV.

FORMAT A4 3125 02/2



* (1262 lb/hr)

STREAM *	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Water	H_2S Rich Gas	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	216	-	1,984	915	-	244	20,638
GAS FLOW MMSCFD	-	2.62	2.35	-	12.60	-	-	1.98	-	-
lb/hr	297,500	3,584	1,376	2,416	3,540	27,769	13,350	7,574	2,733	283,419

CRUDE SOURCE: Canadian Light Sour Blend

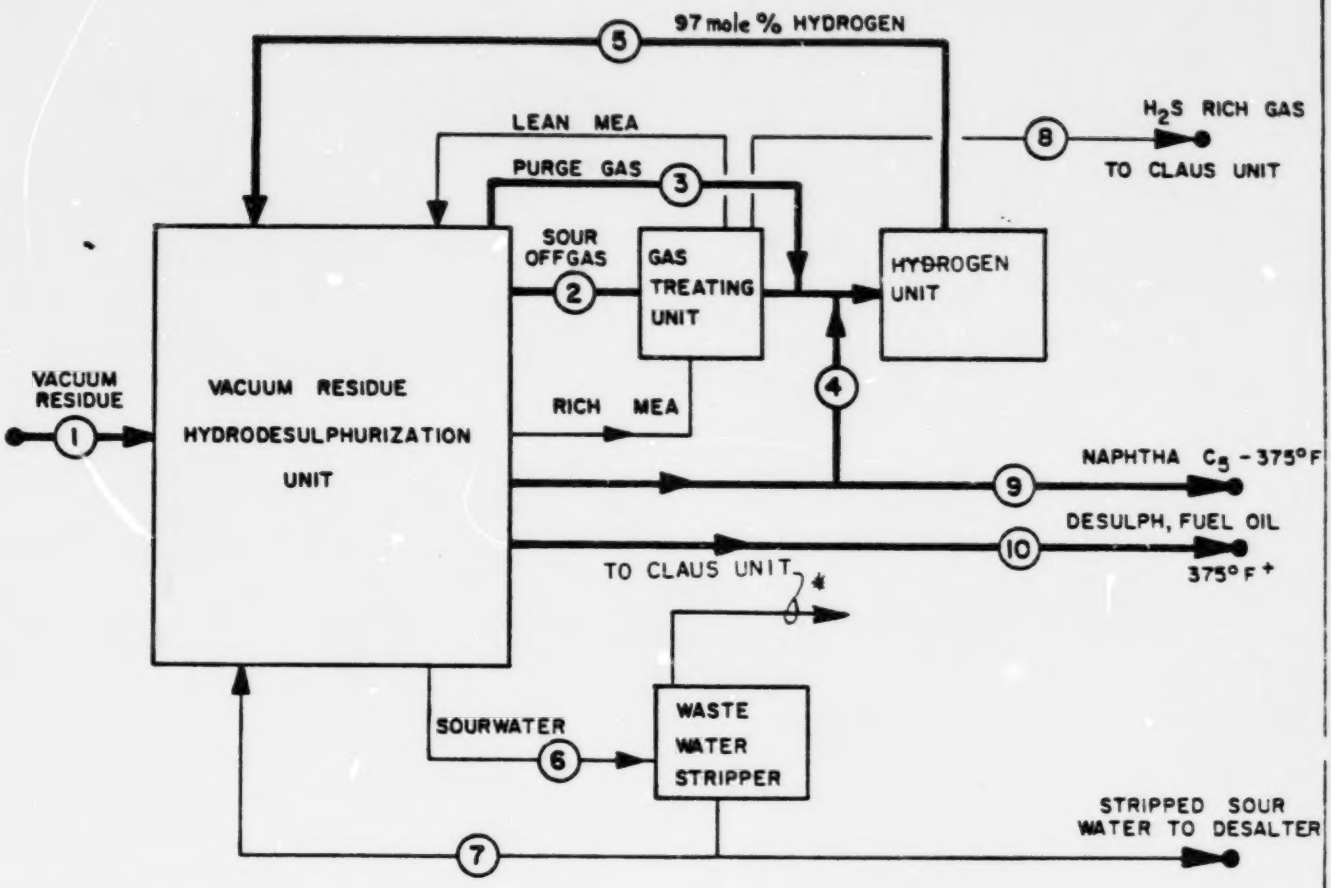
PROCESS: Gulf HDS II

WT %S IN 375+ PRODUCT: 1.0

TABLE 5.2.14

CLIENT ENVIRONMENT CANADA		FAIT/MADE F.V.	VERIFIE/CHKD	DATE	
PROJ. EVALUATION OF DESULPHURIZATION PROCESSES		APPR.	ECH./SCALE	DATE REV.	
MATERIAL BALANCE FOR THE OVERALL COMPLEX		CONTR. / SUBDIV.	ELEMENT	NO.	REV.

FORMAT 3125 02/2



* (1757 lb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Pure Gas	Naphtha to H_2 Unit	Makeup H_2	Sour Water	Stripped Water	H_2S Rich Gas	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	288	-	2,080	1,371	-	360	20,532
GAS FLOW MMSCFD	-	4.52	4.29	-	17.17	-	-	2.37	-	-
lb/hr	297,500	4,240	2,516	3,239	4,822	29,113	20,000	9,072	4,036	279,065

CRUDE SOURCE: Canadian Light Sour Blend

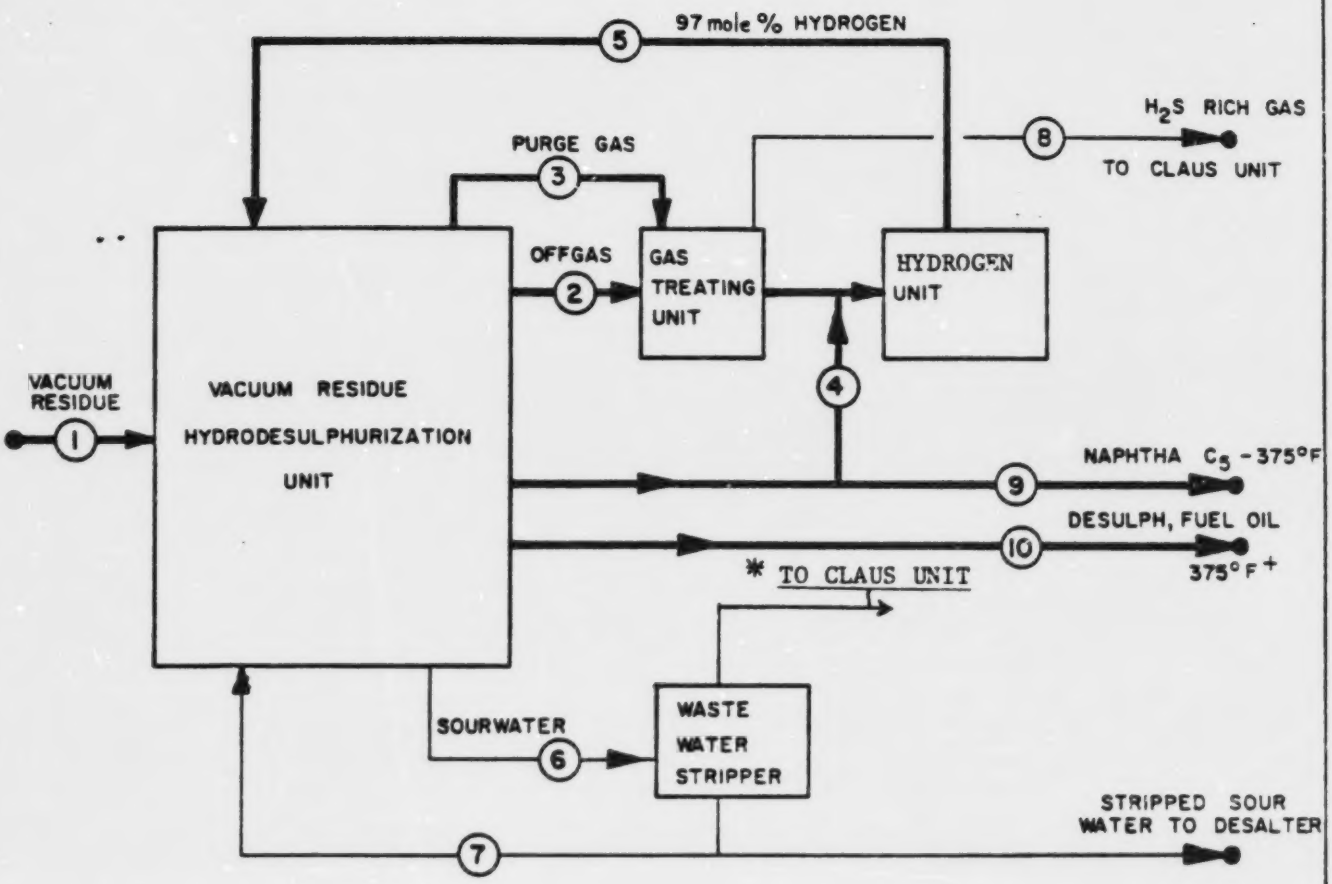
PROCESS: Gulf HDS II

WT %S IN 375+ PRODUCT: 0.5

TABLE 5.2.15

CLIENT: ENVIRONMENT CANADA	FAIT/MADE: F.V.	VERIFIE/CHKD	DATE
PROJ: EVALUATION OF DESULPHURIZATION PROCESSES	APPR.	ECH./SCALE	DATE REV
MATERIAL BALANCE FOR THE OVERALL COMPLEX	CONTR. / SUBDIV.	ELEMENT	NO. / REV

FORMAT 3125 02/2



* (560 lb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit †	Makeup H_2	Sour Water	Stripped S.W. Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	402	-	1164	617	-	-	20,038
GAS FLOW MMSCFD	-	1.41	2.86	-	16.0	-	-	1.63	-	-
lb/hr	297,555	1376	1525	4460	4494	16980	9000	6226	-	287,633

CRUDE SOURCE: Iranian Light

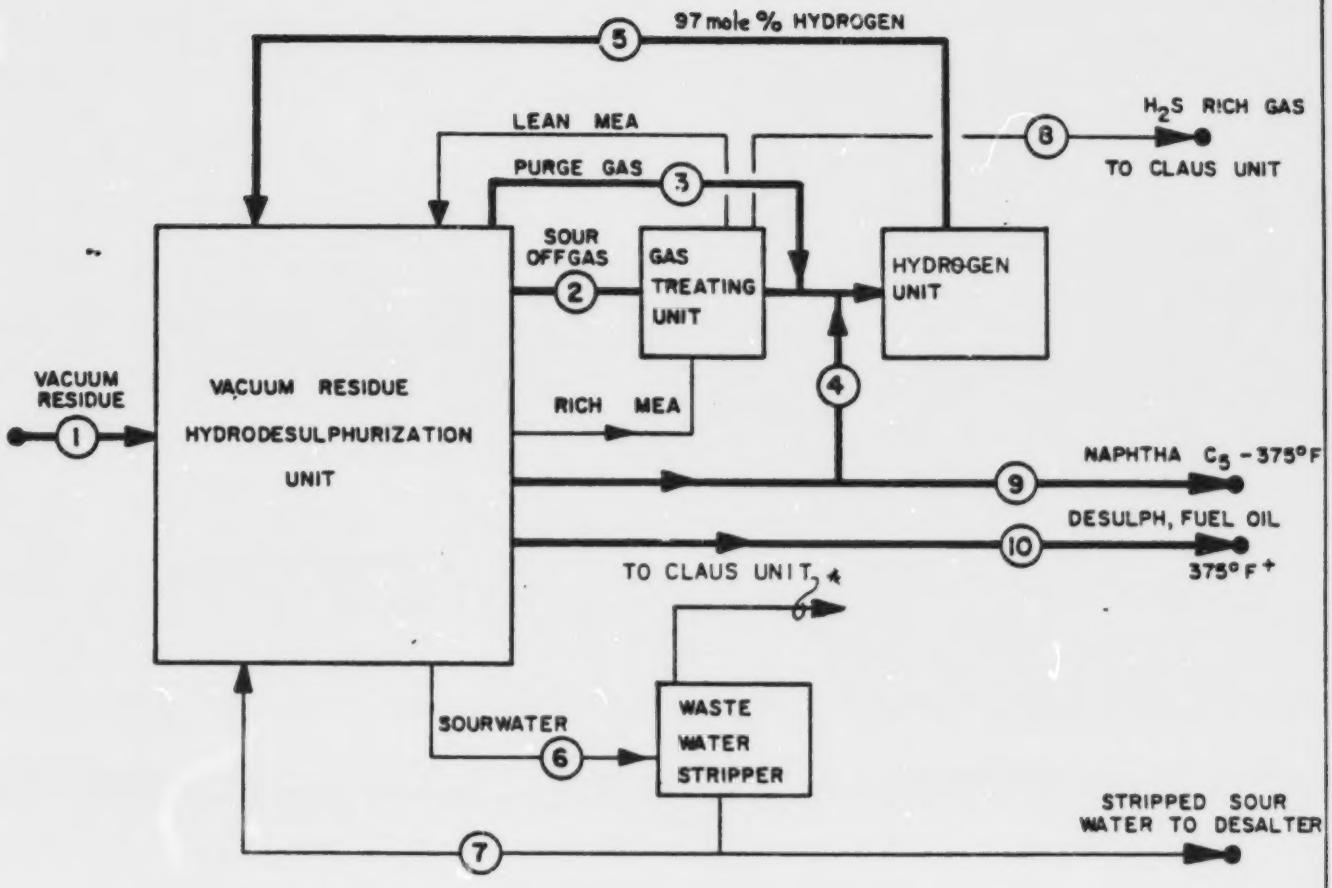
PROCESS: UOP RCD - UNIBON

WT %S IN 375+ PRODUCT: 1.0

† Supplemented by 14 BPSD of 375-650°F cut

TABLE 5.2.16

CLIENT ENVIRONMENT CANADA	FAIT/MADE F.V.	VERIFIE/CHKD	DATE
PROJ EVALUATION OF DESULPHURIZATION PROCESSES	APPR.	ECH./SCALE	DATE REV.
MATERIAL BALANCE FOR THE OVERALL COMPLEX	CONTR. / SUBDIV.	ELEMENT	NO. / REV



* (1320 kb/hr)

STREAM*	1	2	3	4	5	6	7	8	9	10
IDENTIFICATION	Residue	Sour Offgas	Purge Gas	Naphtha to H_2 Unit †	Makeup H_2	Sour Water	Stripped S.W. Recycle	Claus Unit Feed	Excess Naphtha	Desulph. Fuel Oil
LIQUID FLOW BPSD	20,000	-	-	550	-	1,288	617	-	-	19,943
GAS FLOW MMSCFD	-	2.55	5.26	-	25.00	-	-	1.94	-	-
lb/hr	297,500	2,498	2,803	6,161	7,020	17,953	9,000	7,427	-	284,601

CRUDE SOURCE: Canadian Light Sour Blend

PROCESS: UOP RCD - UNIBON

WT %S IN 375+ PRODUCT: 1.0

† Supplemented by 42 BPSD of 375-650°F cut

TABLE 5.2.17

CLIENT ENVIRONMENT CANADA	FAIT/MADE F.V.	VERIFIE/CHKD	DATE
PROJ EVALUATION OF DESULPHURIZATION PROCESSES	APPR.	ECH./SCALE	DATE REV.
MATERIAL BALANCE FOR THE	CONTR. / SUBDIV.	ELEMENT	NO. / REV.

FORMAT 302/2

Table 5.2.18 SUMMARY MATERIAL BALANCES

STREAM	1	2	3	4	5	6	7	8	9	10
Iranian Light residue, 1.0 wt %S case										
GULF	297,555	3523	1742	3085	4118	41,098	40,000	5816	1,568	285,876
CHEVRON	297,555	3638	3697	1640	5393	19,686	11,667	6111	2,447	284,724
UOP	297,555	2498	2803	6161	7020	17,953	9,000		-	
H-Oil	297,555	15,148		-	5620	28,017	20,000	6050	13,480	273,750
Iranian Light residue, 0.5 wt %S case										
GULF	297,555	3777	2618	4055	5360			7359	2,487	282,313
CHEVRON	297,555	3893	5538	5260	7864	31,825	23,680	8634	500	281,718
H-Oil	297,555	19,399		-	7020	28,360	20,000	7913	18,148	266,004
Canadian Light Sour Blend residue, 1.0 wt %S case										
GULF	297,500	3584	1376	2416	3540	27,769	13,350	7574	2,733	283,419
CHEVRON	297,500	3492	4953	1545	6040	30,301	21,680	7366	3,744	282,167
UOP	297,500	2498	2803	6161	7020	17,953	9,000	7427	-	284,601
H-Oil	297,500	16,889		-	5620	24,564	16,000	7400	15,170	269,832
Canadian Light Sour Blend residue, 0.5 wt %S case										
GULF	297,500	4240	2516	3239	4822	29,113	20,000	9072	4,036	279,065
CHEVRON	297,500	3967	6910	3460	8710	52,817	43,355	8557	4,137	278,067
H-Oil	297,500	23,328		-	7527	23,900	16,000	8956	25,585	255,850
Tia Juana residue, 1.0 wt %S case										
H-Oil	298,800	16,334		2700	6432	28,192	20,000	6614	14,032	271,310
Tia Juana residue, 0.5 wt %S case										
H-Oil	298,800	22,864		-	8990	28,398	20,000	8201	28,385	253,680

All flows in lb/hr

	Tia Juana		Iranian Light		Canadian Light Sour Blend	
wt %S in 375°F⁻	1.0	0.5	1.0	0.5	1.0	0.5
a) Gulf HDS						
C_1-C_4 wt % charge	-	-	0.65	0.82	0.56	0.79
C_5- 375°F vol% charge	-	-	2.32	3.04	2.44	3.47
375°F⁺ vol% charge	-	-	101.99	101.89	103.19	102.66
Total C_5^+ vol% charge			104.31	104.93	105.63	106.13
b) Chevron VRDS						
C_1-C_4 wt % charge	-	-	1.19	1.63	1.50	2.06
C_5-375°F vol% charge	-	-	1.90	2.70	2.50	3.60
375°F⁺ vol% charge	-	-	99.99	99.40	99.60	98.90
Total C_5^+ vol% charge			101.89	102.10	102.10	102.50
c) H-Oil						
C_1-C_4 wt % charge	2.20	4.30	5.40	5.80	5.20	7.00
C_5-375°F vol% charge	8.27	13.98	6.50	8.60	7.20	11.90
375°F⁺ vol% charge	96.40	92.10	96.60	95.30	96.50	93.50
Total C_5^+ vol% charge	104.67	106.08	103.10	103.90	103.70	105.40
d) RCD Unibon						
C_1-C_4 wt % charge	-	-	0.20	-	0.33	-
C_5-375°F vol% charge	-	-	2.11	-	3.14	-
375°F⁺ vol% charge	-	-	100.26	-	100.00	-
Total C_5^+ vol% charge			102.37		103.14	

Table 5.2.19 COMPARISON OF PRODUCTS YIELDS

	H-OIL Tia Juana 1.0 wt%S.	H-OIL Tia Juana 0.5 wt%S.
1. POWER KW/B. of DFO		
a) HDS Unit (incl. Booster Comp.)	8.85	11.72
b) H_2 Unit	3.80	5.64
c) MEA + SWS Units	0.39	0.41
	13.04	17.77
2. STEAM Mlb/B. of DFO		
a) HDS Unit	0.0093	0.0098
b) H_2 Unit	(0.0529)	(0.0782)
c) MEA - SWS Units	0.0390	0.0467
	(0.0046)	(0.0217)
3. FUEL MMBTU/B. of DFO		
a) HDS Unit	0.0950	0.1216
b) H_2 Unit	0.2178	0.3040
	0.3128	0.4256

DFO : Desulphurized Fuel Oil

Table 5.2.20a SUMMARY OF MAJOR UTILITIES - TIA JUANA 1.0 & 0.5 wt% S. CASES

	H-OIL	GULF	CHEVRON	UOP
1. POWER KW/B. of DFO				
a) HDS Unit (incl. Booster Comp.)	7.58	7.21	3.14	4.30
b) H_2 Unit	3.53	2.04	2.96	2.08
c) MEA + SWS Units	0.39	0.36	0.35	0.40
	11.50	9.61	6.45	6.78
2. STEAM Mlb/B. of DFO				
a) HDS Unit	0.0088	0.0145	(0.0040)	0.0104
b) H_2 Unit	(0.0487)	(0.0284)	(0.0416)	(0.0480)
c) MEA + SWS Units	0.0340	0.0082	0.0332	0.0330
	(0.0059)	(0.0054)	(0.0124)	(0.0046)
3. FUEL MMBTU/B. of DFO				
a) HDS Unit	0.1004	0.0372	0.0420	0.0280
b) H_2 Unit	0.1858	0.1270	0.1680	0.1370
	0.2862	0.1642	0.2100	0.1650

DFO : Desulphurized Fuel Oil

Table 5.2.20b SUMMARY OF MAJOR UTILITIES - IRANIAN LIGHT 1.0 wt% S. CASES

	H-OIL	GULF	CHEVRON	EXXON
1. POWER KW/B. of DFO				
a) HDS Unit (incl. Booster Comp.)	8.92	6.22	4.66	5.10
b) H_2 Unit	4.48	2.83	3.37	2.00
c) MEA + SWS Units	0.41	0.39	0.31	0.40
	13.81	9.44	8.34	7.50
2. STEAM Mlb/B. of DFO				
a) HDS Unit	0.0089	0.0196	(0.0060)	(0.0630)
b) H_2 Unit	(0.0619)	(0.0265)	(0.0460)	(0.0370)
c) MEA + SWS Units	0.0460	0.0127	0.0480	0.0330
	(0.0070)	0.0058	(0.0040)	(0.0670)
3. FUEL MMBTU/B. of DFO				
a) HDS Unit	0.0101	0.0383	0.0420	0.0500
b) H_2 Unit	0.2420	0.1700	0.2030	0.1560
	0.2521	0.2083	0.2450	0.2060

DFO : Desulphurized Fuel Oil

Table 5.2.20c SUMMARY OF MAJOR UTILITIES - IRANIAN LIGHT 0.5 wt% S. Cases

	H-OIL	GULF	CHEVRON
1. POWER KW/B. of DFO			
a) HDS Unit (incl. Booster Compressor)	6.88	5.32	3.55
b) H_2 Unit	3.66	1.41	2.62
c) MEA - SWS Units	0.39	0.35	0.30
	10.93	7.08	6.47
2. STEAM Mlb/B. of DFO			
a) HDS Unit	0.0062	0.0194	(0.0120)
b) H_2 Unit	(0.0487)	(0.0263)	(0.0472)
c) MEA - SWS Units	0.0415	0.0058	0.0400
	(0.0010)	(0.0011)	(0.0192)
3. FUEL MMBTU/B. of DFO			
a) HDS Unit	0.0249	0.0378	0.0420
b) H_2 Unit	0.1907	0.1115	0.1870
	0.2156	0.1493	0.2290

DFO : Desulphurized Fuel Oil

Table 5.2.20d SUMMARY OF MAJOR UTILITIES - CANADIAN LIGHT SOUR BLEND 1.0 wt% S. CASES

	H-OIL	GULF	CHEVRON
1. POWER KW/B. of DFO			
a) HDS Unit (incl. Booster Comp.)	8.80	5.60	5.60
b) H_2 Unit	4.89	1.95	3.80
c) MEA - SWS Units	0.43	0.34	0.30
	14.12	7.89	9.70
2. STEAM Mlb/B. of DFO			
a) HDS Unit	0.0064	0.0242	(0.0105)
b) H_2 Unit	(0.0674)	(0.0360)	(0.0667)
c) MEA - SWS Units	0.0503	0.0074	0.0538
	(0.0107)	(0.0044)	(0.0234)
3. FUEL MMBTU/B. of DFO			
a) HDS Unit	0.0717	0.0379	0.0425
b) H_2 Unit	0.2630	0.1520	0.2679
	0.3347	0.1919	0.3104

DFO : Desulphurized Fuel Oil

Table 5.2.20e SUMMARY OF MAJOR UTILITIES - CANADIAN LIGHT SOUR BLEND 0.5 wt% S. CASES

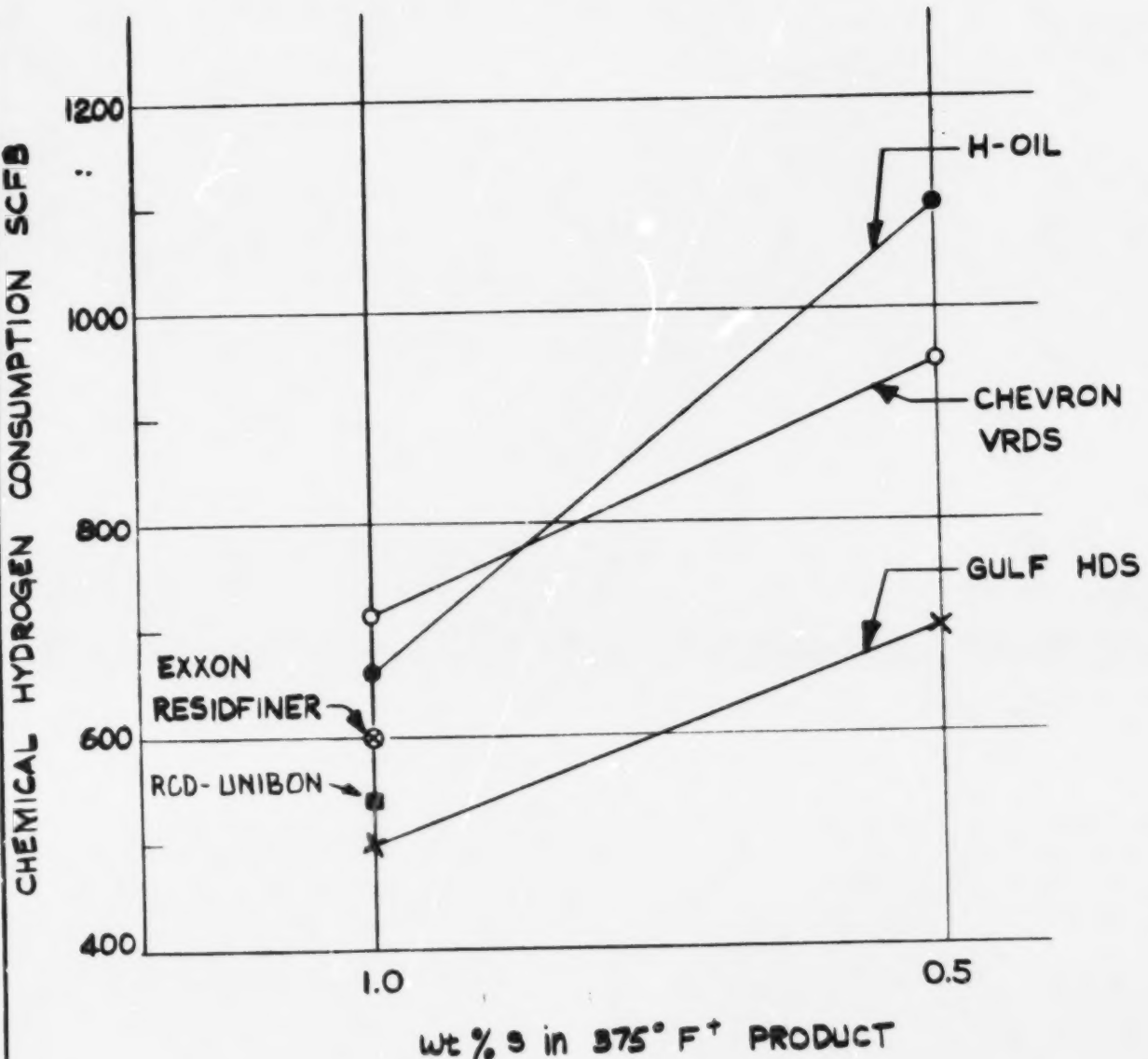


FIGURE 5-2-1

CLIENT ENVIRONMENT CANADA EVALUATION OF DESULPH. PROCESSES		FAIT MADE F.V.	VERIFIE/CHKD PSK	DATE
PROJ.		APPR.	ECH SCALE	DATE REV
COMPARISON OF CHEMICAL HYDROGEN CONSUMPTION AGHAJARI VACUUM RESIDUE		CONTR. / SUBDIV.	ELEMENT	NO. / REV

CANADIAN LIGHT SOURBLEND RESIDUE

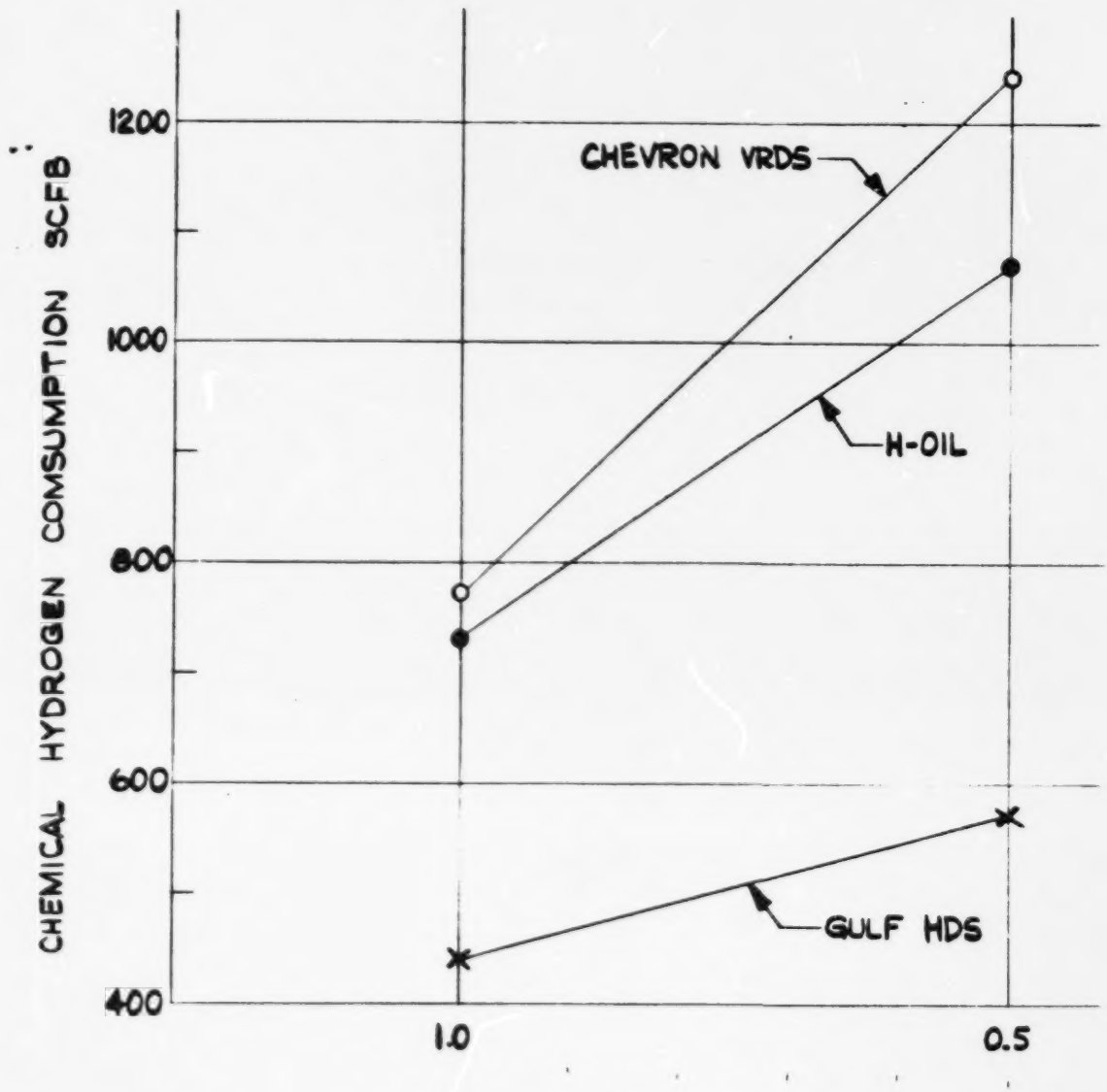


FIGURE 5-2-2

CLIENT ENVIRONMENT CANADA EVALUATION OF DESULPH. PROCESSES		FAIT/MADE F.V.	VERIFIE/CHKD PSK	DATE
PROJ.		APPR.	ECH/SCALE	DATE REV
COMPARISON OF CHEMICAL HYDROGEN CONSUMPTION: CANADIAN LIGHT SOURBLEND VACUUM RESIDUE		CONTR / SUBDIV.	ELEMENT	NO. / REV

5. TECHNICAL/ECONOMIC ASSESSMENT (cont'd)

5.3 Investment and Operating Costs

The process investment battery limit costs including the service facilities for the HDS, hydrogen, gas treating and sour water treating units are presented in Figures 5.3.1 to 5.3.4. The installed costs were derived for 20,000 BPSD with Edmonton as the reference location. The equipment included in deriving the investment costs are shown in Figures 5.1.1 to 5.1.4. The costs for the 10,000-30,000 BPSD capacity range were derived using exponents shown in Figures 5.3.1 to 5.3.4. The exponents were calculated from licensor data. It should be noted that the reactors are installed in multiple trains and therefore, the common exponent of 0.6 is appliable only to the product recovery section.

Tables 5.3.5 to 5.3.16 provide details of the total production costs for processing a barrel of desulphurized fuel oil for each residue case and sulphur levels of 0.5 and 1.0 wt % in the 375°F$^+$ product. They include the capital and operating costs related to the hydrogen, gas treating and sour water treating units as well. A net byproduct credit for sulphur has been assigned after taking due consideration of the processing costs through the Claus sulphur recovery unit. The basis for calculating the capital charges and operating costs are provided in Appendix A. As explained in Section 2.5 the capital charges have been derived for two types of financing to highlight the sensitivity of unit production costs with respect to the capital charges.

Figures 5.3.5 to 5.3.11 present the unit costs of processing various residua in $/Bbl of 375°F$^+$ product for the two extreme levels of capital charges. The costs for processing to 0.5 wt %S level are in general 10-20% higher than for processing to 1.0 wt %S in product. Most of this increase stems from higher fuel consumption and catalyst costs. There is a substantial difference in the unit costs between the fixed bed and ebullated bed (H-Oil) processes, the former being considerably higher. In order to study the differences, the cost data for the Iranian Light 1.0 wt %S case have been broken down in terms of capital charges (utility type financing), variable costs, catalyst costs and net byproduct credits and presented in Figure 5.3.5, an examination of which shows the following:

- The capital charges for H-Oil are considerably lower than those of any of the fixed bed processes - $0.89/Bbl for H-Oil compared to $1.23 for Chevron VRDS and $1.10 for Gulf HDS. This can be attributed to the following factors: HRI claim that a single train of two reactors can process up to 20,000-28,000 BPSD residue whereas three parallel trains are required by Chevron VRDS for 30,000 BPSD. H-Oil does not require scrubbers for the removal of H_2S and NH_3 in the high pressure gas circuit. The gas treatment is performed at a lower pressure resulting in considerable savings. However, this is partially offset by the increased reactor total operating pressure necessary to maintain the hydrogen partial pressure. It appears that this is not too great a disadvantage. HRI have included a

5. TECHNICAL/ECONOMIC ASSESSMENT (cont'd)

5.3 Investment and Operating Costs (cont'd)

vacuum tower to generate a recycle gas oil stream to the reactor but claim this is not necessary.

- The fixed and variable operating costs do not vary significantly; \$1.16/Bbl for H-Oil, \$1.03/Bbl for Chevron VRDS and \$1.15/Bbl for Gulf HDS. The small differences are due to the fixed operating costs which are investment related and utility consumptions.

- The catalyst costs constitute a significant portion of the total production costs and vary appreciably amongst the processes considered. The catalyst costs for the H_2 unit are small in relation to the HDS unit. Therefore the costs shown in Figure 5.3.5 can be directly compared as HDS catalyst costs. Within the fixed bed processes themselves, Chevron VRDS has \$0.87/Bbl catalyst costs compared to \$0.30/Bbl for Gulf HDS. But for this differential (\$0.57/Bbl) the processing costs using Chevron VRDS and Gulf HDS are almost equal. In comparison, H-Oil has much lower catalyst cost at \$0.15/Bbl.

- The other major difference between H-Oil and the fixed bed processes is due to the net byproduct credit assigned. In the case of H-Oil, the net yield of light gases is sufficient to generate the hydrogen makeup required. Therefore all of the naphtha (C_5-375°F) produced commands a premium value. Moreover slightly lower investment and operating costs for the H_2 units result for all of the H-Oil cases studied due to approximately 10% higher investment costs associated with naphtha feedstock rather than all gas feedstock for equivalent hydrogen production. If unit costs for makeup hydrogen and fuel value for the light gases were to be assumed the economics might be expected to change with the unit cost per barrel of finished fuel nearing \$2.00 for H-Oil for the case illustrated in Figure 5.3.5.

Table 5.3.1 : COST SUMMARY

PROCESS : H-OIL CRUDE : TIA JUANA wt%s in FUEL OIL : 1.0

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	21.070		35.360		48.122	
ii) H_2 UNIT	12.750		20.000		25.900	
iii) GAS TREATING & SWS UNITS	3.870		6.080		8.010	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.207		0.414		0.621	
D) WORKING CAPITAL (WC)	3.120		5.066		6.765	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.147		1.860		2.490	
TAXES & INSURANCE	0.754		1.229		1.641	
GENERAL SALES AND ADMINISTRATION	0.565		0.922		1.230	
B) VARIABLE COSTS						
POWER	0.524		0.992		1.382	
FUEL GAS	1.528		3.056		4.584	
STEAM	(0.036)		(0.072)		(0.108)	
BOILERFEED WATER	0.016		0.033		0.049	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	0.650		1.257		1.872	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(2.262)		(4.525)		(6.786)	
CREDIT FOR SULPHUR	0.122		0.244		0.366	
TOTAL OPERATING COSTS (MM$/yr)	3.241		5.552		6.960	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.03		0.83		0.74	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.261	-	3.686	-	4.922
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	3.840	5.819	6.479	9.435	8.645	12.590
WC	0.310	0.310	0.503	0.503	0.672	0.672
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	2.34	3.71	1.94	2.92	1.72	2.67

Table 5.3.2 : COST SUMMARY

PROCESS : H-OIL CRUDE : TIA JUANA wt% in FUEL OIL : 0.5

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
1) HDS UNIT	33.802		55.477		75.520	
11) H_2 UNIT	14.300		22.500		29.290	
111) GAS TREATING & SWS UNITS	4.247		6.690		8.665	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.249		0.496		0.745	
D) WORKING CAPITAL (WC)	4.216		6.972		9.294	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.557		2.562		3.432	
TAXES & INSURANCE	1.049		1.723		2.307	
GENERAL SALES AND ADMINISTRATION	0.787		1.292		1.730	
B) VARIABLE COSTS						
POWER	0.694		1.290		1.785	
FUEL GAS	1.983		3.966		5.949	
STEAM	(0.181)		(0.361)		(0.542)	
BOILERFEED WATER	0.007		0.013		0.020	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	1.413		2.773		4.041	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(5.010)		(10.020)		(15.030)	
CREDIT FOR SULPHUR	0.162		0.324		0.486	
TOTAL OPERATING COSTS (MM$/yr)	2.694		3.799		4.418	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	0.90		0.63		0.49	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	3.081	-	5.080	-	6.808
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	5.402	7.867	8.872	12.919	11.687	17.020
WC	0.420	0.420	0.697	0.697	0.929	0.929
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	2.81	4.66	2.21	3.73	1.90	3.24

Table 5.3.3 : COST SUMMARY

PROCESS : H-OIL CRUDE : IRANIAN LIGHT wt% in FUEL OIL : 1.0

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	15.770		26.530		36.105	
ii) H_2 UNIT	11.520		18.000		21.000	
iii) GAS TREATING & SWS UNITS	4.350		5.240		7.020	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.135		0.270		0.405	
D) WORKING CAPITAL (WC)	2.631		4.120		5.314	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	0.966		1.515		1.952	
TAXES & INSURANCE	0.633		0.995		1.283	
GENERAL SALES AND ADMINISTRATION	0.475		0.747		0.962	
B) VARIABLE COSTS						
POWER	0.475		0.876		1.205	
FUEL GAS	1.422		2.843		4.265	
STEAM	0.066		0.131		0.197	
BOILERFEED WATER	0.007		0.013		0.020	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	0.483		0.953		1.409	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(2.312)		(4.624)		(6.936)	
CREDIT FOR SULPHUR	0.110		0.220		0.330	
TOTAL OPERATING COSTS (MM$/yr)	2.558		3.306		4.927	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	0.81		0.62		0.52	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	1.898	-	2.986	-	3.848
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	3.372	4.911	5.277	7.685	6.801	9.904
WC	0.263	0.263	0.412	0.412	0.531	0.531
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	1.97	3.07	1.53	2.40	1.30	2.03

Table 5.3.4 : COST SUMMARY

PROCESS : H-Oil CRUDE : IRANIAN LIGHT wt% in FUEL OIL : 0.5

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	20.040		33.900		46.135	
ii) H_2 UNIT	12.299		19.300		25.120	
iii) GAS TREATING & SWS UNITS	3.795		5.768		7.446	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.225		0.450		0.675	
D) WORKING CAPITAL (WC)	2.997		4.870		6.502	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.100		1.791		2.390	
TAXES & INSURANCE	0.745		1.209		1.613	
GENERAL SALES AND ADMINISTRATION	0.559		0.906		1.210	
B) VARIABLE COSTS						
POWER	0.519		1.038		1.557	
FUEL GAS	1.659		3.319		4.978	
STEAM	(0.058)		(0.117)		(0.175)	
BOILERFEED WATER	0.013		0.026		0.039	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	0.569		1.137		1.705	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(3.104)		(6.208)		(9.312)	
CREDIT FOR SULPHUR	0.131		0.262		0.393	
TOTAL OPERATING COSTS (MM$/yr)	2.456		3.600		4.638	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	0.79		0.58		0.50	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.168	-	3.538	-	4.722
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	3.835	5.585	6.224	9.064	8.302	12.090
WC	0.300	0.300	0.487	0.487	0.650	0.650
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	2.11	3.37	1.65	2.68	1.45	2.36

Table 5.3.5 : COST SUMMARY

PROCESS : H-OIL CRUDE : CDN.LT. SOUR BLEND wt% in FUEL OIL : 1.0

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	12.642		20.580		27.335	
ii) H_2 UNIT	10.390		16.300		21.000	
iii) GAS TREATING & SWS UNITS	4.050		6.280		8.174	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.215		0.431		0.646	
D) WORKING CAPITAL (WC)	2.272		3.604		4.724	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	0.829		1.314		1.724	
TAXES & INSURANCE	0.563		0.668		1.168	
GENERAL SALES AND ADMINISTRATION	0.423		0.501		0.876	
B) VARIABLE COSTS						
POWER	0.443		0.822		1.147	
FUEL GAS	1.270		2.540		3.810	
STEAM	(0.016)		(0.033)		(0.049)	
BOILERFEED WATER	0.018		0.026		0.039	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	0.565		1.104		1.616	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(2.549)		(5.098)		(7.647)	
CREDIT FOR SULPHUR	0.142		0.284		0.426	
TOTAL OPERATING COSTS (MM$/yr)	1.921		2.365		3.320	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	0.61		0.37		0.36	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	1.625	-	2.584	-	3.391
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	2.903	4.227	4.585	6.678	6.017	8.761
WC	0.227	0.227	0.360	0.360	0.472	0.472
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	1.60	2.54	1.15	1.90	1.05	1.70

Table 5.3.6 : COST SUMMARY

PROCESS : H-OIL CRUDE :CDN LT SOUR BLEND wt% in FUEL OIL : 0.5

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	17.703		29.360		39.562	
ii) H_2 UNIT	12.870		20.200		25.800	
iii) GAS TREATING & SWS UNITS	3.894		5.889		8.565	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.540		1.081		1.621	
D) WORKING CAPITAL (WC)	2.884		4.639		6.119	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.038		1.685		2.218	
TAXES & INSURANCE	0.702		1.138		1.497	
GENERAL SALES AND ADMINISTRATION	0.527		0.854		1.123	
B) VARIABLE COSTS						
POWER	0.555		1.044		1.459	
FUEL GAS	1.583		3.167		4.750	
STEAM	(0.053)		(0.105)		(0.158)	
BOILERFEED WATER	0.013		0.026		0.039	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	0.854		1.643		2.414	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(4.25)		(8.50)		(12.75)	
CREDIT FOR SULPHUR	0.166		0.332		0.498	
TOTAL OPERATING COSTS (MM$/yr)	1.368		1.521		1.330	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	0.45		0.25		0.15	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.042	-	3.327	-	4.378
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	4.122	5.106	5.711	8.317	7.517	10.946
WC	0.288	0.288	0.464	0.464	0.612	0.612
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	1.89	2.88	1.26	2.23	1.04	1.89

Table 5.3.7 : COST SUMMARY

PROCESS : GULF HDS CRUDE : IRANIAN LIGHT wt%s in FUEL OIL : 1.0

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	31.374		49.797		64.810	
ii) H_2 UNIT	9.686		15.199		20.187	
iii) GAS TREATING & SWS UNITS	2.173		3.223		4.099	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.702		1.170		1.580	
D) WORKING CAPITAL (WC)	3.632		5.668		7.406	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.325		2.069		2.702	
TAXES & INSURANCE	0.894		1.364		1.820	
GENERAL SALES AND ADMINISTRATION	0.671		1.023		1.365	
B) VARIABLE COSTS						
POWER	0.434		0.802		1.084	
FUEL GAS	0.845		1.690		2.535	
STEAM	(0.072)		(0.144)		(0.216)	
BOILERFEED WATER	0.007		0.013		0.020	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	1.146		2.089		2.661	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.257)		(0.513)		(0.770)	
CREDIT FOR SULPHUR	0.107		0.214		0.321	
TOTAL OPERATING COSTS (MM$/yr)	5.333		8.844		11.762	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.55		1.29		1.14	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.620	-	4.093	-	5.346
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	4.491	6.539	7.026	10.233	9.174	13.360
WC	0.363	0.363	0.567	0.567	0.741	0.741
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	2.96	4.33	2.39	3.46	2.10	3.03

Table 5.3.8 : COST SUMMARY

PROCESS : GULF HDS CRUDE : IRANIAN LIGHT wt% in FUEL OIL : 0.5

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	34.317		53.850		69.860	
ii) H_2 UNIT	11.348		17.807		23.651	
iii) GAS TREATING & SWS UNITS	1.998		2.890		3.657	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.697		1.394		1.830	
D) WORKING CAPITAL (WC)	3.957		6.192		8.072	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.446		2.259		2.943	
TAXES & INSURANCE	0.953		1.520		1.981	
GENERAL SALES AND ADMINISTRATION	0.715		1.140		1.486	
B) VARIABLE COSTS						
POWER	0.409		0.758		1.029	
FUEL GAS	0.870		1.740		2.610	
STEAM	(0.050)		(0.108)		(0.158)	
BOILERFEED WATER	0.013		0.026		0.039	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	1.353		2.411		3.223	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.401)		(0.802)		(1.203)	
CREDIT FOR SULPHUR	0.130		0.260		0.390	
TOTAL OPERATING COSTS (MM$/yr)	5.671		9.441		12.58	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.69		1.41		1.25	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.860	-	4.473	-	5.830
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	5.023	7.314	7.830	11.402	10.204	14.860
WC	0.396	0.396	0.619	0.619	0.807	0.807
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	3.31	4.87	2.67	3.88	2.35	3.40

Table 5.3.9 : COST SUMMARY

PROCESS : GULF HDS CRUDE : CDN LT SOUR BLEND wt%s in FUEL OIL : 1.0

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	21.616		33.920		43.328	
ii) H_2 UNIT	9.040		14.186		18.464	
iii) GAS TREATING & SWS UNITS	2.133		3.170		4.049	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.291		0.564		0.780	
D) WORKING CAPITAL (WC)	2.734		4.264		5.482	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.101		1.560		1.884	
TAXES & INSURANCE	0.678		1.055		1.275	
GENERAL SALES AND ADMINISTRATION	0.509		0.791		0.956	
B) VARIABLE COSTS						
POWER	0.288		0.576		0.864	
FUEL GAS	0.779		1.558		2.337	
STEAM	(0.009)		(0.018)		(0.027)	
BOILERFEED WATER	0.007		0.013		0.020	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	0.446		0.848		1.104	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.441)		(0.882)		(1.322)	
CREDIT FOR SULPHUR	0.140		0.280		0.420	
TOTAL OPERATING COSTS (MM$/yr)	3.751		6.018		7.751	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.11		0.89		0.76	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	1.967	-	3.077	-	3.710
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	3.490	5.083	5.432	7.910	6.595	9.561
WC	0.273	0.273	0.426	0.426	0.548	0.548
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	2.22	3.28	1.75	2.58	1.46	2.13

Table 5.3.10 : COST SUMMARY

PROCESS : GULF HDS CRUDE : CDN LT SOUR BLEND wt% in FUEL OIL : 0.5

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	27.478		43.417		56.509	
ii) H_2 UNIT	10.541		16.541		21.529	
iii) GAS TREATING & SWS UNITS	2.352		3.504		4.508	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.339		0.678		0.916	
D) WORKING CAPITAL (WC)	3.345		5.248		6.829	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.229		1.926		2.505	
TAXES & INSURANCE	0.830		1.298		1.689	
GENERAL SALES AND ADMINISTRATION	0.623		0.974		1.267	
B) VARIABLE COSTS						
POWER	0.320		0.640		0.960	
FUEL GAS	0.986		1.973		2.959	
STEAM	(0.042)		(0.084)		(0.126)	
BOILERFEED WATER	0.007		0.013		0.020	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	0.637		1.156		1.478	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.650)		(1.300)		(1.950)	
CREDIT FOR SULPHUR	0.160		0.330		0.490	
TOTAL OPERATING COSTS (MM$/yr)	4.333		7.163		9.532	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.28		1.06		0.94	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.423	-	3.808	-	4.953
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	4.273	6.222	6.686	9.738	8.698	12.667
WC	0.334	0.334	0.525	0.525	0.683	0.683
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	2.65	3.96	2.13	3.16	1.87	2.76

Table 5.3.11 : COST SUMMARY

PROCESS : CHEVRON VRDS CRUDE : IRANIAN LIGHT wt% in FUEL OIL : 1.0

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	32.196		50.520		65.413	
ii) H_2 UNIT	10.960		17.200		22.680	
iii) GAS TREATING & SWS UNITS	1.912		2.816		3.620	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	2.250		4.400		6.600	
D) WORKING CAPITAL (WC)	3.873		6.100		8.017	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.369		2.138		2.780	
TAXES & INSURANCE	0.921		1.440		1.872	
GENERAL SALES AND ADMINISTRATION	0.691		1.080		1.404	
B) VARIABLE COSTS						
POWER	0.483		0.894		1.242	
FUEL GAS	1.061		2.122		3.183	
STEAM	(0.154)		(0.309)		(0.463)	
BOILERFEED WATER	0.007		0.013		0.020	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	2.857		5.714		8.572	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.385)		(0.770)		(1.155)	
CREDIT FOR SULPHUR	0.134		0.268		0.402	
TOTAL OPERATING COSTS (MM$/yr)	7.217		12.747		18.097	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	2.19		1.94		1.84	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.704	-	4.232	-	5.503
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	4.755	6.925	7.416	10.799	9.643	14.042
WC	0.387	0.387	0.610	0.610	0.802	0.802
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	3.76	5.25	3.17	4.33	2.90	3.91

Table 5.3.12 : COST SUMMARY

PROCESS : CHEVRON CRUDE : IRANIAN LIGHT wt%s in FUEL OIL : 0.5

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
1) HDS UNIT	36.510		57.290		74.480	
11) H_2 UNIT	13.890		21.800		28.374	
111) GAS TREATING & SWS UNITS	1.830		2.720		3.410	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	3.878		5.267		7.900	
D) WORKING CAPITAL (WC)	4.577		7.083		9.285	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.583		2.476		3.216	
TAXES & INSURANCE	1.067		1.665		2.163	
GENERAL SALES AND ADMINISTRATION	0.800		1.249		1.622	
B) VARIABLE COSTS						
POWER	0.327		0.653		0.980	
FUEL GAS	1.238		2.477		3.715	
STEAM	(0.036)		(0.072)		(0.108)	
BOILERFEED WATER	0.023		0.046		0.069	
COOLING WATER	0.013		0.026		0.039	
C) CATALYSTS & CHEMICALS	3.335		6.671		10.005	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.080)		(0.160)		(0.240)	
CREDIT FOR SULPHUR	0.162		0.324		0.486	
TOTAL OPERATING COSTS (MM$/yr)	8.662		15.585		22.177	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	2.65		2.38		2.26	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	3.134	-	4.909	-	6.376
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	5.493	8.000	8.576	12.491	11.141	16.225
WC	0.458	0.458	0.708	0.708	0.929	0.929
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	4.47	6.21	3.81	5.16	3.50	4.67

Table 5.3.13 : COST SUMMARY

PROCESS : CHEVRON CRUDE : CND LT SOUR BLEND wt%s in FUEL OIL : 1.0

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	25.024		39.268		50.160	
ii) H_2 UNIT	11.917		18.700		24.340	
iii) GAS TREATING & SWS UNITS	1.738		2.540		3.265	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	1.630		3.170		4.740	
D) WORKING CAPITAL (WC)	3.312		5.211		6.883	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.177		1.837		2.361	
TAXES & INSURANCE	0.796		1.239		1.593	
GENERAL SALES AND ADMINISTRATION	0.597		0.929		1.195	
B) VARIABLE COSTS						
POWER	0.255		0.509		0.764	
FUEL GAS	1.158		2.317		3.475	
STEAM	(0.172)		(0.344)		(0.516)	
BOILERFEED WATER	0.010		0.020		0.030	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	1.669		3.042		4.510	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.600)		(1.200)		(1.800)	
CREDIT FOR SULPHUR	0.139		0.278		0.416	
TOTAL OPERATING COSTS (MM$/yr)	5.262		8.864		12.268	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.61		1.39		1.24	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.321	-	3.630	-	4.666
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	4.098	5.967	6.382	9.295	8.213	11.959
WC	0.331	0.331	0.521	0.521	0.688	0.688
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	2.96	4.25	2.48	3.41	2.16	3.02

Table 5.3.14 : COST SUMMARY

PROCESS : CHEVRON VRDS; CRUDE : CDN LT SOUR BLEND; wt% in FUEL OIL : 0.5

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	41.005		64.450		83.790	
ii) H_2 UNIT	14.845		23.300		30.326	
iii) GAS TREATING & SWS UNITS	2.350		2.600		3.290	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	2.623		5.247		7.870	
D) WORKING CAPITAL (WC)	4.905		7.764		10.174	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.745		2.732		3.551	
TAXES & INSURANCE	1.174		1.836		2.386	
GENERAL SALES AND ADMINISTRATION	0.881		1.377		1.790	
B) VARIABLE COSTS						
POWER	0.378		0.757		1.135	
FUEL GAS	1.555		3.110		4.665	
STEAM	(0.215)		(0.430)		(0.645)	
BOILERFEED WATER	0.030		0.059		0.089	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	2.550		5.100		7.550	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.661)		(1.322)		(1.984)	
CREDIT FOR SULPHUR	0.161		0.322		0.483	
TOTAL OPERATING COSTS (MM$/yr)	7.831		13.778		19.260	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	2.41		2.12		1.97	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	3.456	-	5.421	-	7.044
- OBL	-	0.055	-	0.073	-	0.095
INTEREST - IBL - OBL	6.046	8.805	9.457	13.772	12.288	17.896
WC	0.490	0.490	0.776	0.776	1.017	1.017
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	4.42	6.35	3.69	5.20	3.34	4.65

Table 5.3.15 : COST SUMMARY

PROCESS : UOP-RCD UNIBON CRUDE : IRANIAN LIGHT wt% in FUEL OIL : 1.0

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
1) HDS UNIT	25.018		43.560		60.250	
ii) H_2 UNIT	19.946		15.600		20.300	
iii) GAS TREATING & SWS UNITS	2.172		3.220		4.204	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	1.139		2.278		3.417	
D) WORKING CAPITAL (WC)	3.150		5.289		7.206	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.026		0.026		0.026	
MAINTENANCE	1.153		1.929		2.622	
TAXES & INSURANCE	0.764		1.277		1.733	
GENERAL SALES AND ADMINISTRATION	0.574		0.958		1.300	
B) VARIABLE COSTS						
POWER	0.280		0.534		0.761	
FUEL GAS	0.835		1.670		2.505	
STEAM	(0.042)		(0.083)		(0.125)	
BOILERFEED WATER	0.008		0.016		0.024	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	2.042		4.084		6.126	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	-		-		-	
CREDIT FOR SULPHUR	0.134		0.268		0.402	
TOTAL OPERATING COSTS (MM$/yr)	5.981		10.890		15.588	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.82		1.65		1.58	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.228	-	3.743	-	5.085
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	3.938	5.735	6.576	9.576	8.925	12.998
WC	0.315	0.315	0.528	0.528	0.721	0.721
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	3.11	4.35	2.73	3.77	2.56	3.49

Table 5.3.16 : COST SUMMARY

PROCESS : UOP-RCD UNIBON CRUDE : CAN SOUR BLEND wt% in FUEL OIL : 1.0

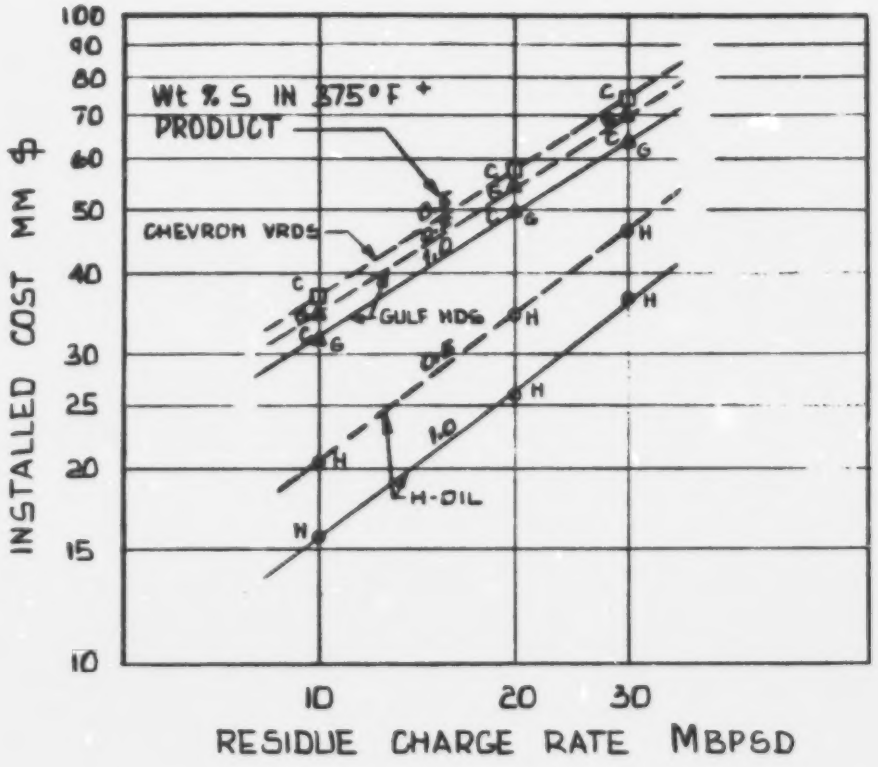
CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	30.220		52.632		72.800	
ii) H_2 UNIT	13.120		20.500		26.500	
iii) GAS TREATING & SWS UNITS	2.172		3.220		4.204	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	0.967		1.934		2.901	
D) WORKING CAPITAL (WC)	3.806		6.380		8.664	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.046		0.046		0.046	
MAINTENANCE	1.382		2.312		3.134	
TAXES & INSURANCE	0.932		1.556		2.108	
GENERAL SALES AND ADMINISTRATION	0.699		1.167		1.581	
B) VARIABLE COSTS						
POWER	0.322		0.644		0.966	
FUEL GAS	1.244		2.448		3.672	
STEAM	(0.042)		(0.083)		(0.125)	
BOILERFEED WATER	0.008		0.016		0.024	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	0.887		1.774		2.661	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	-		-		-	
CREDIT FOR SULPHUR	0.140		0.280		0.420	
TOTAL OPERATING COSTS (MM$/yr)	5.785		10.351		14.681	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.76		1.58		1.49	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.730	-	4.581	-	6.210
- OBL	-	0.044	-	0.058	-	0.134
INTEREST - IBL - OBL	4.801	6.992	8.015	11.672	10.857	15.811
WC	0.381	0.381	0.638	0.638	0.866	0.866
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	3.35	4.86	2.90	4.16	2.68	3.83

Table 5.3.17 : COST SUMMARY

PROCESS : EXXON CRUDE : IRANIAN LIGHT wt% in FUEL OIL : 0.6

CAPACITY BPSD	10,000		20,000		30,000	
I CAPITAL INVESTMENT (MM$)						
A) INSIDE BATTERY LIMITS (IBL)						
i) HDS UNIT	30.320		47.620		61.927	
ii) H_2 UNIT	9.686		16.541		21.529	
iii) GAS TREATING & SWS UNITS	2.173		3.190		4.099	
B) OUTSIDE BATTERY LIMITS (OBL)	1.100		1.460		1.900	
C) INITIAL CATALYST CHARGE	1.935		3.870		5.805	
D) WORKING CAPITAL (WC)	3.617		5.814		7.621	
II OPERATING COSTS (MM$/yr)						
A) FIXED COSTS						
DIRECT OPERATING LABOUR	0.184		0.184		0.184	
SUPERVISION	0.026		0.026		0.026	
MAINTENANCE	1.282		2.042		2.656	
TAXES & INSURANCE	0.866		1.376		1.790	
GENERAL SALES AND ADMINISTRATION	0.649		1.032		1.343	
B) VARIABLE COSTS						
POWER	0.364		0.694		0.989	
FUEL GAS	1.042		2.084		3.126	
STEAM	(0.592)		(1.184)		(1.775)	
BOILERFEED WATER	0.010		0.020		0.030	
COOLING WATER	0.003		0.007		0.010	
C) CATALYSTS & CHEMICALS	1.873		3.745		5.618	
D) BYPRODUCTS						
CREDIT FOR NAPHTHA	(0.524)		(1.049)		(1.573)	
CREDIT FOR SULPHUR	0.158		0.306		0.464	
TOTAL OPERATING COSTS (MM$/yr)	5.361		9.303		12.443	
OPERATING COST $/Bbl DESULPHURIZED FUEL OIL	1.61		1.40		1.25	
TYPE OF FINANCING	UTILITY	EQUITY	UTILITY	EQUITY	UTILITY	EQUITY
III CAPITAL CHARGES (MM$/yr)						
DEPRECIATION - IBL	-	2.530	-	4.041		5.253
- OBL	-	0.044	-	0.058	-	0.076
INTEREST - IBL - OBL	4.460	6.495	7.086	10.320	9.219	13.428
WC	0.362	0.362	0.581	0.581	0.762	0.762
PRODUCTION COST $/Bbl DESULPHURIZED FUEL OIL	3.06	4.45	2.55	3.66	2.25	3.21

ALBERTA LOCATION
2ND QUARTER 1977
ACCURACY ± 20%
IRANIAN LIGHT RESIDUE



EXPONENT

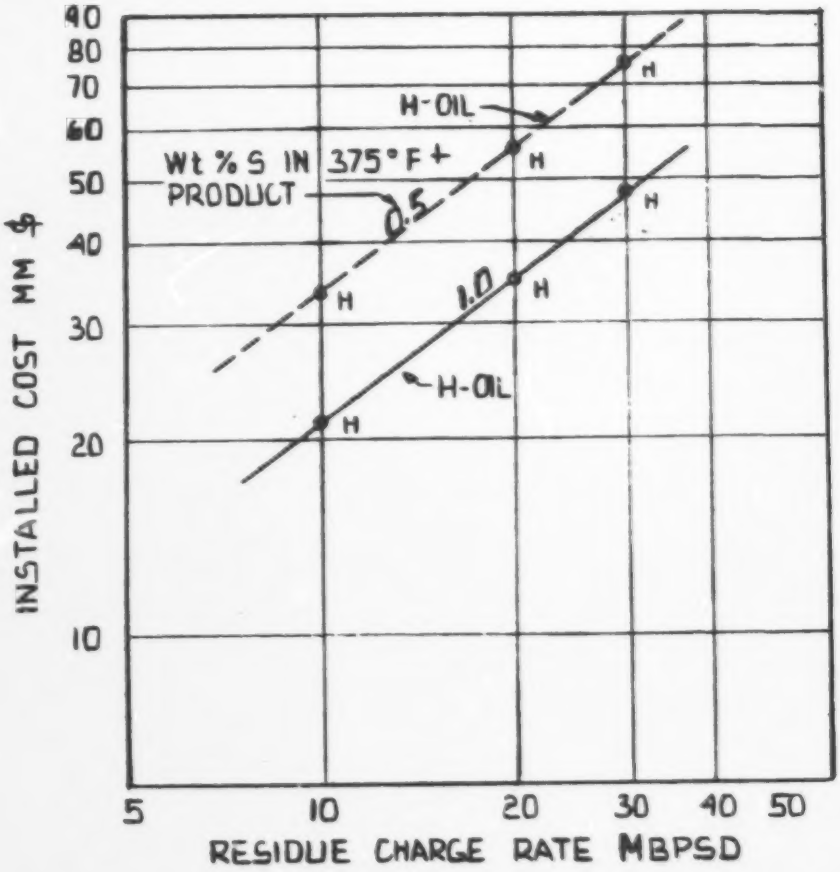
□ C	= CHEVRON VRDS	0.65
▲ G	= GULF HDS	0.65
⊙ H	= H-OIL	0.76

FIGURE 5-3-1a

CLIENT ENVIRONMENT CANADA	FAIT/MADE N.M.	VERIFIE/CHKD	DATE	
PROJ. EVALUATION OF DESULPH. PROCESSES	APPR.	ECH./SCALE —	DATE REV.	
INSTALLED COSTS FOR RESIDUE HDS UNIT IRANIAN LIGHT RESIDUE	CONTR. / SUBDIV.	ELEMENT	NO.	REV.

FORMAT "A" 3128-02/3

ALBERTA LOCATION
2ND QUARTER 1977
ACCURACY ± 20 %
TIA JUANA RESIDUE



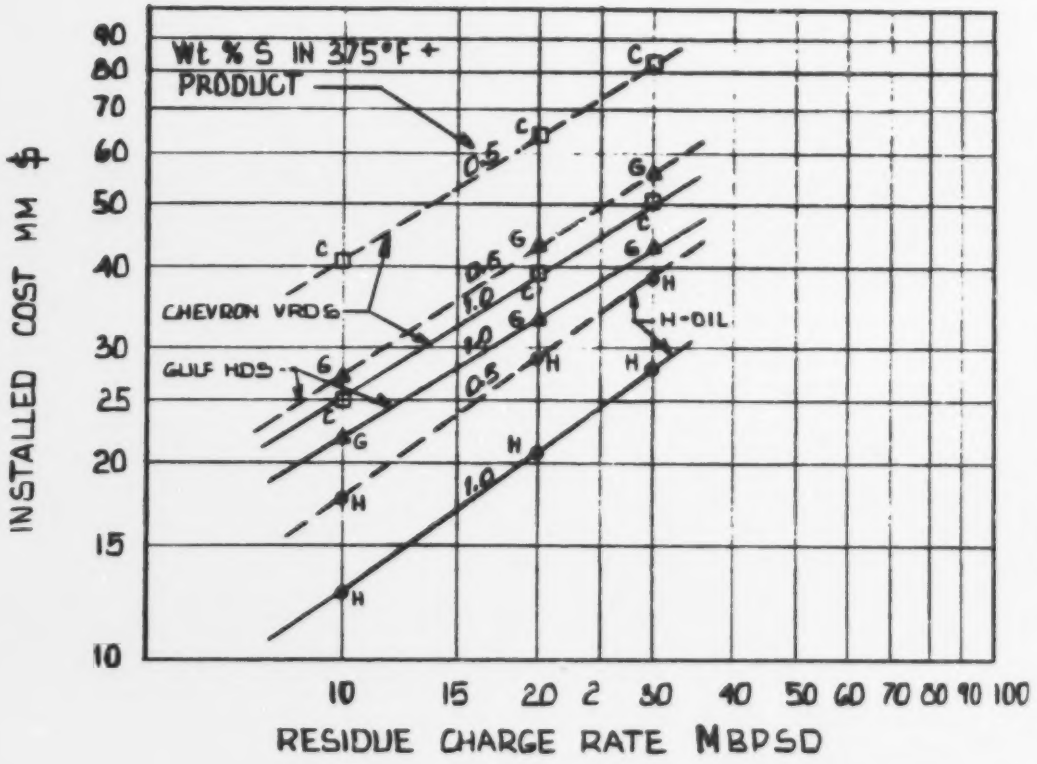
EXPONENT
⊙H = H-OIL 0.76

FIGURE 5-3-1b

CLIENT ENVIRONMENT CANADA	FAIT/MADE N.M.	VERIFIE/CHKD	DATE	
PROJ. EVALUATION OF DEGULPH. PROCESSES	APPR.	ECH./SCALE —	DATE REV.	
INSTALLED COSTS FOR RESIDUE HDS UNIT TIA JUANA RESIDUE	CONTR.	SUBDIV.	ELEMENT	NO. REV

FORMAT "A"
3126-02/2

ALBERTA LOCATION
2ND QUARTER 1977
ACCURACY ± 20%
LIGHT SOURBLEND RESIDUE



EXPONENT

□ C	= CHEVRON VRDS	0.65
▲ G	= GULF HDS	0.65
⊙ H	= H-OIL	0.74

FIGURE 5-3-1c

CLIENT ENVIRONMENT CANADA	FAIT/MADE N. N.		VERIFIE/CHKD	DATE	
PROJ. EVALUATION OF DESULPH. PROCESSES	APPR.		ECH./SCALE —	DATE REV.	
INSTALLED COSTS FOR RESIDUE HDS UNIT LIGHT SOURBLEND RESIDUE	CONTR.	SUBDIV.	ELEMENT	NO.	REV

FORMAT "A"

ALBERTA LOCATION
2ND QUARTER 1977
97 MOLE % 280 psig AT B.L.
FIGURE 5-1-2 SHOWS
EQUIPMENT INCLUDED.

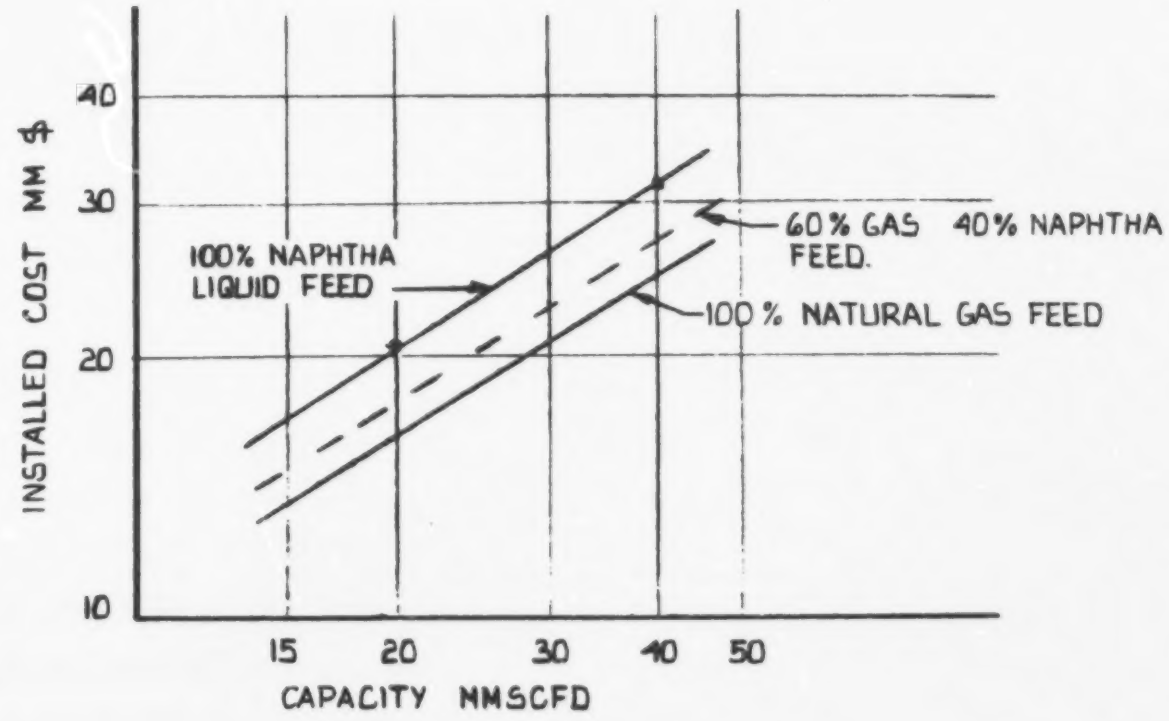


FIGURE 5-3-2

CLIENT ENVIRONMENT CANADA		FAIT/MADE N.M.	VERIFIE/CHKD PSK	DATE
PROJ. EVALUATION OF DESULPH. PROCESSES		APPR.	ECH./SCALE —	DATE REV.
INSTALLED COSTS FOR H_2 UNIT	CONTR. SUBDIV.	ELEMENT	NO.	REV.

FORMAT "A"

ALBERTA LOCATION
2ND QUARTER 1977
ACCURACY ± 20%
FIGURE 5-1-3 SHOWS
EQUIPMENT INCLUDED.

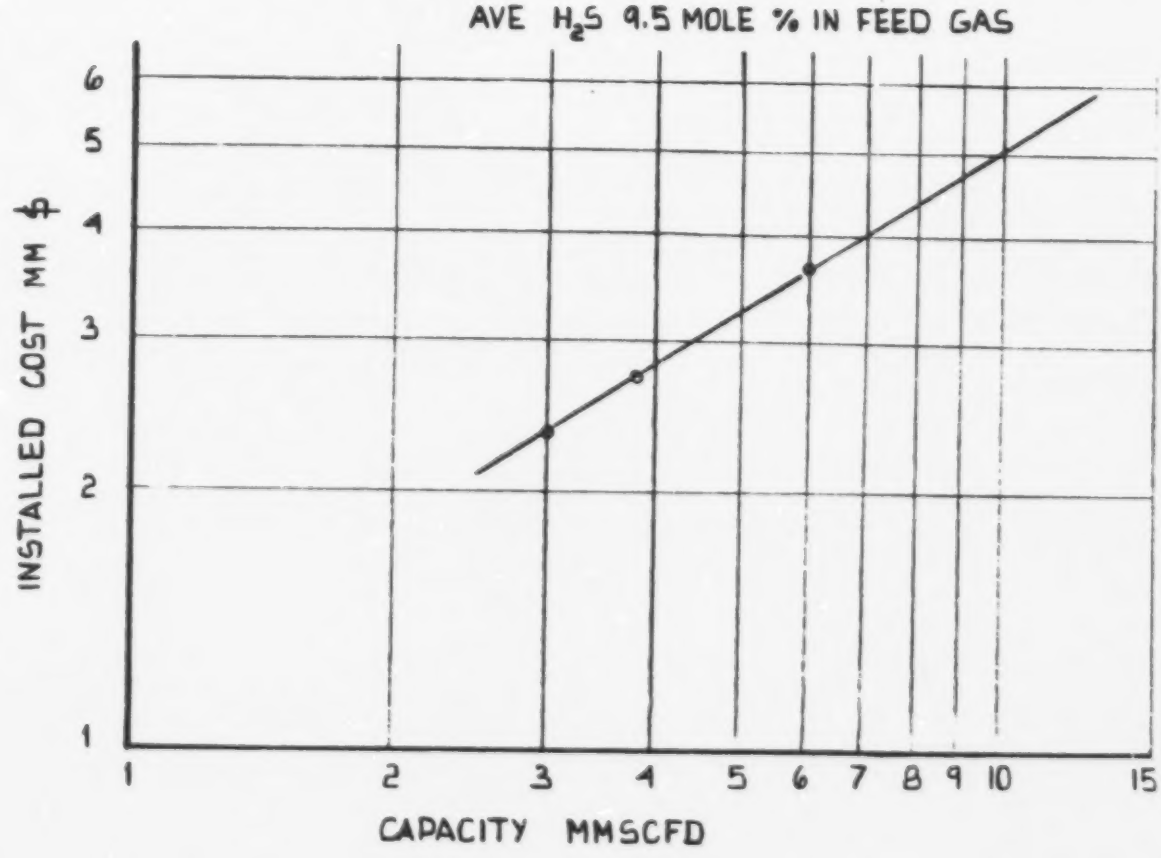


FIGURE 5-3-3

CLIENT ENVIRONMENT CANADA PROJ. EVALUATION OF DESULPH. PROCESSES		FAIT/MADE N. M.	VERIFIE/CHKD PSK	DATE
		APPR.	ECH./SCALE —	DATE REV
INSTALLED COSTS FOR GAS TREATING UNIT.		CONTR. / SUBDIV.	ELEMENT	NO. / REV

FORMAT "A"
2125-02/2

BASIS : ALBERTA LOCATION
DESIGN : SINGLE TOWER / NON REFLUXED
ACCURACY ± 20 %
FIGURE 5-1-4 SHOWS EQUIPMENT INCLUDED.

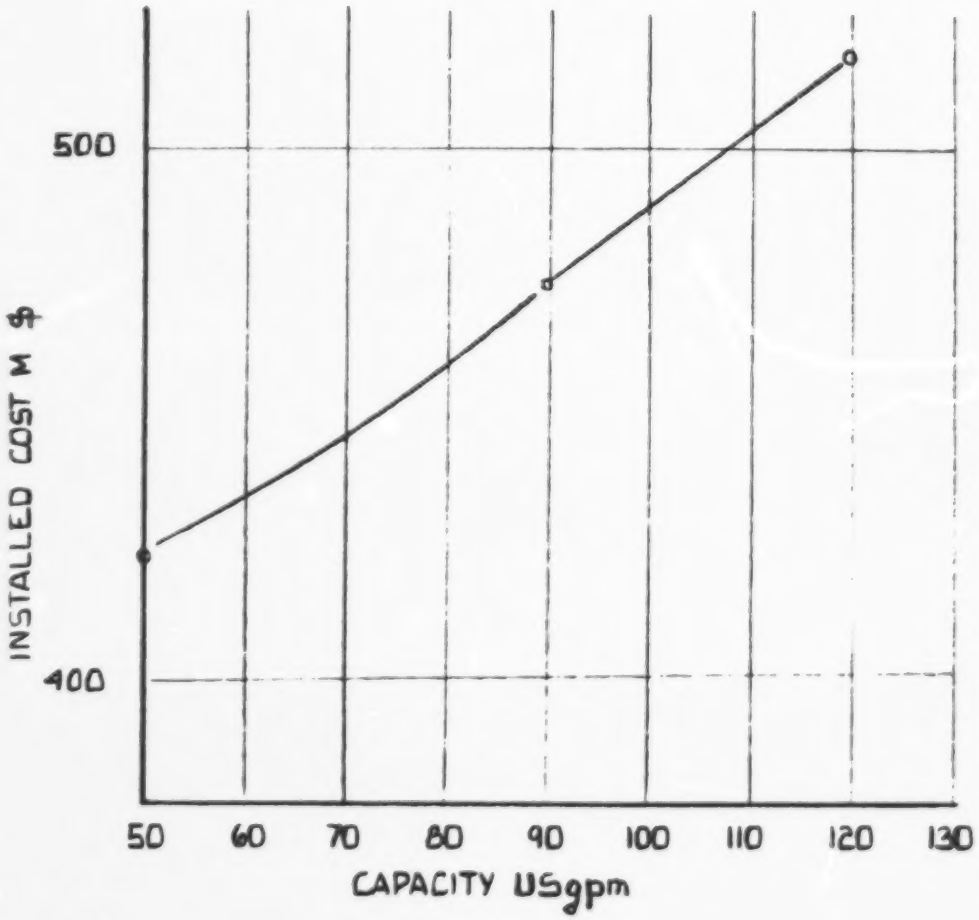


FIGURE 5-3-4

CLIENT ENVIRONMENT CANADA		FAIT/MADE N.M.		VERIFIE/CHKD PSK	DATE
PROJ. EVALUATION OF DESULPH. PROCESSES		APPR.		ECH./SCALE —	DATE REV.
INSTALLED COST OF S.W. STRIPPER UNIT		CONTR.	SUBDIV.	ELEMENT	NO. / REV.

FORMAT A
3125-92/2

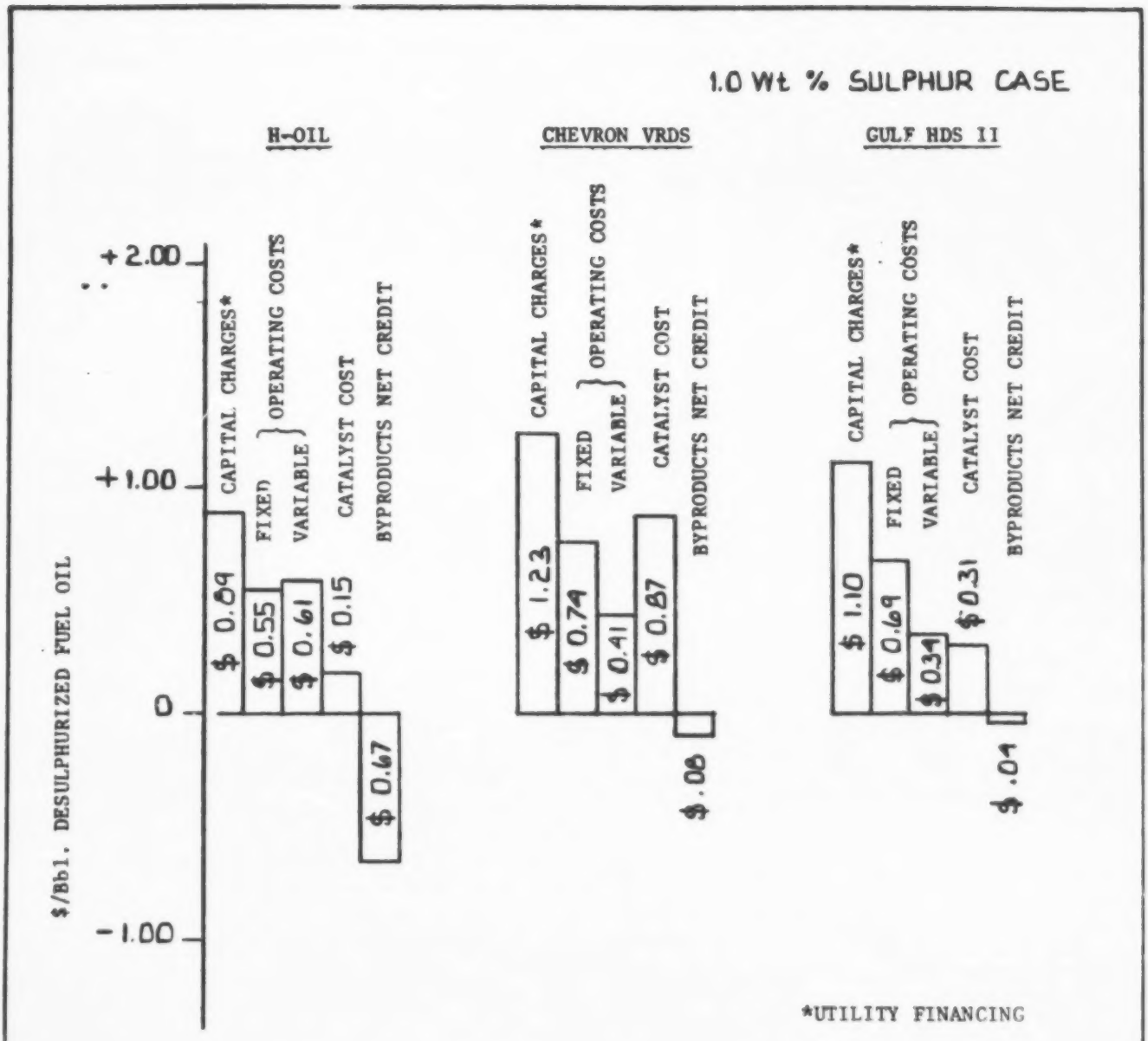


FIGURE 5-3-5

CLIENT ENVIRONMENT CANADA	FAIT/MADE N. M.	VERIFIE/CHKD	DATE
PROJ. EVALUATION OF DESULPH. PROCESSES	APPR.	ECH./SCALE	DATE REV
BREAKDOWN OF COSTS FOR IRANIAN LIGHT	CONTR. / SUBDIV.	ELEMENT	NO.

FORMAT "A"

TIA JUANA

1.0 Wt % S. CASE

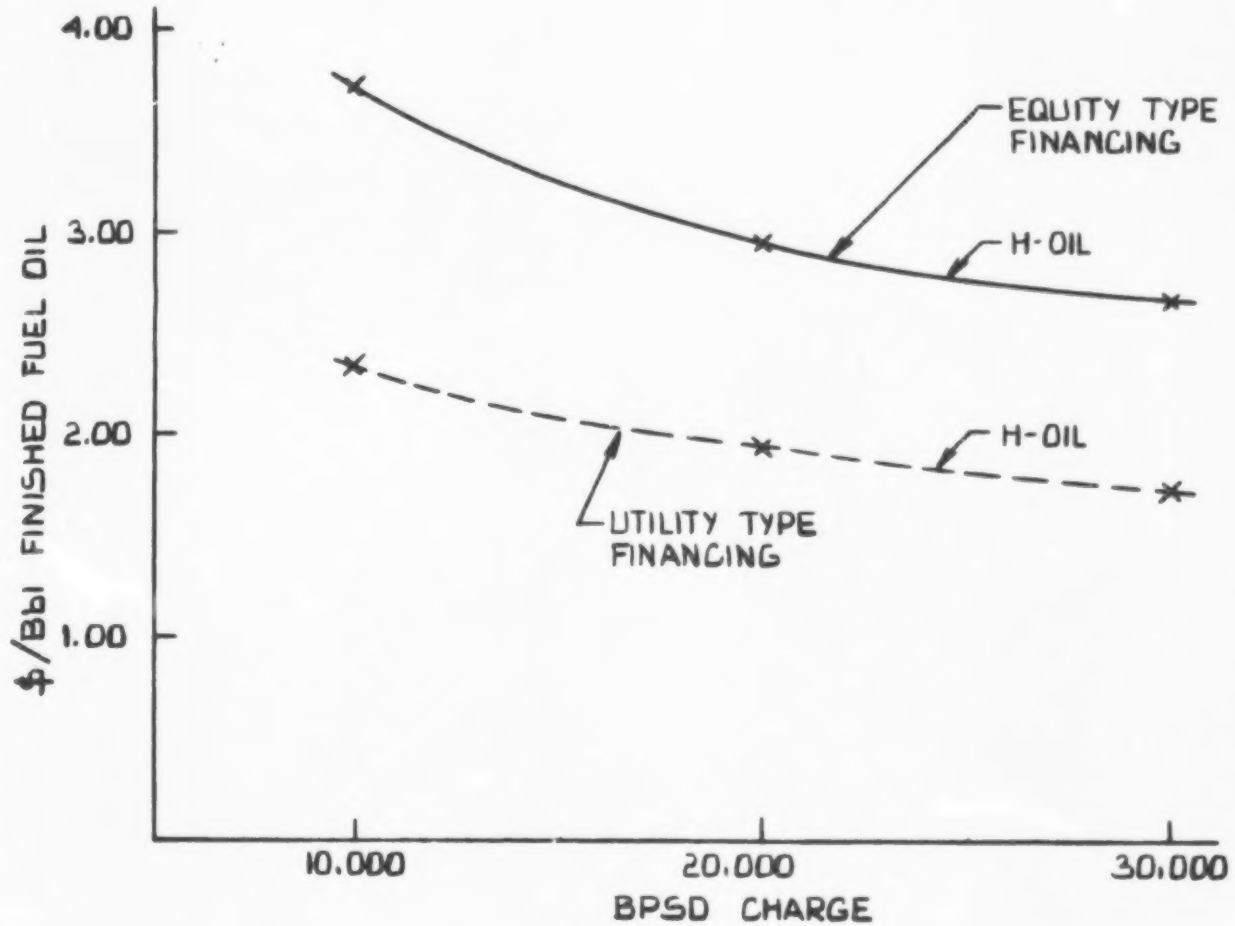


FIGURE 5-3-6

CLIENT	ENVIRONMENT CANADA	DRAWN/MADE N.M.	DATE
PROJ.	EVALUATION OF DESULPH. PROCESSES	APPR.	DATE REV.
	COST OF DESULPHURIZATION	CONTR. SUBDIV.	NO
	TIA JUANA RESIDUE		

FORMAT "A"

3128-02/2

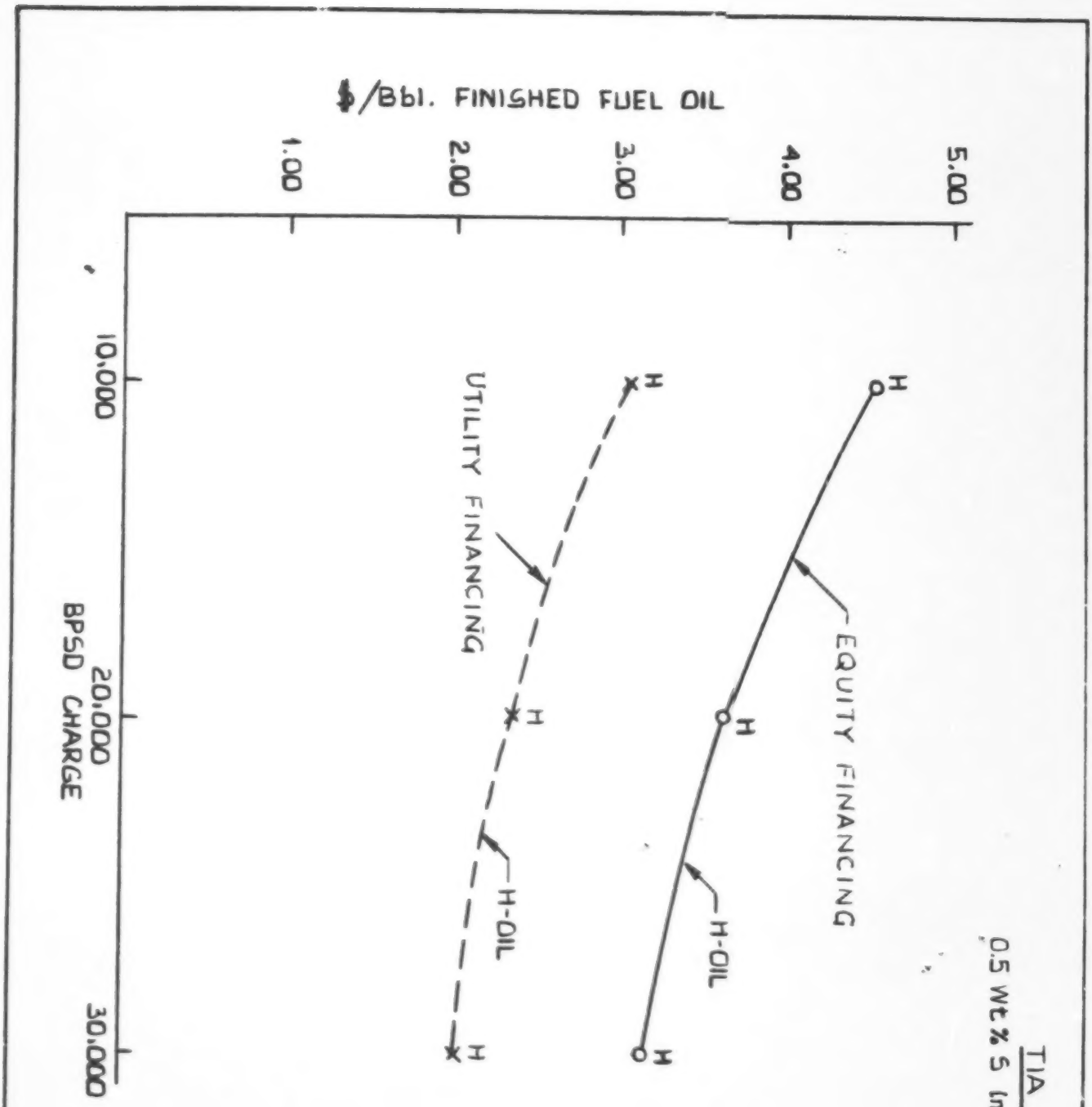


FIGURE 5-3-7

CLIENT ENVIRONMENT CANADA		FAIT/MADE N.M.	VERIFIE/CHKD PSK	DATE
PROJ. EVALUATION OF DESULPH. PROCESSES		APPR.	ECH./SCALE —	DATE REV.
COST OF DESULPHURIZATION TIA JUANA RESIDUE		CONTR. SUBDIV.	ELEMENT	NO. REV

FORMAT "A"

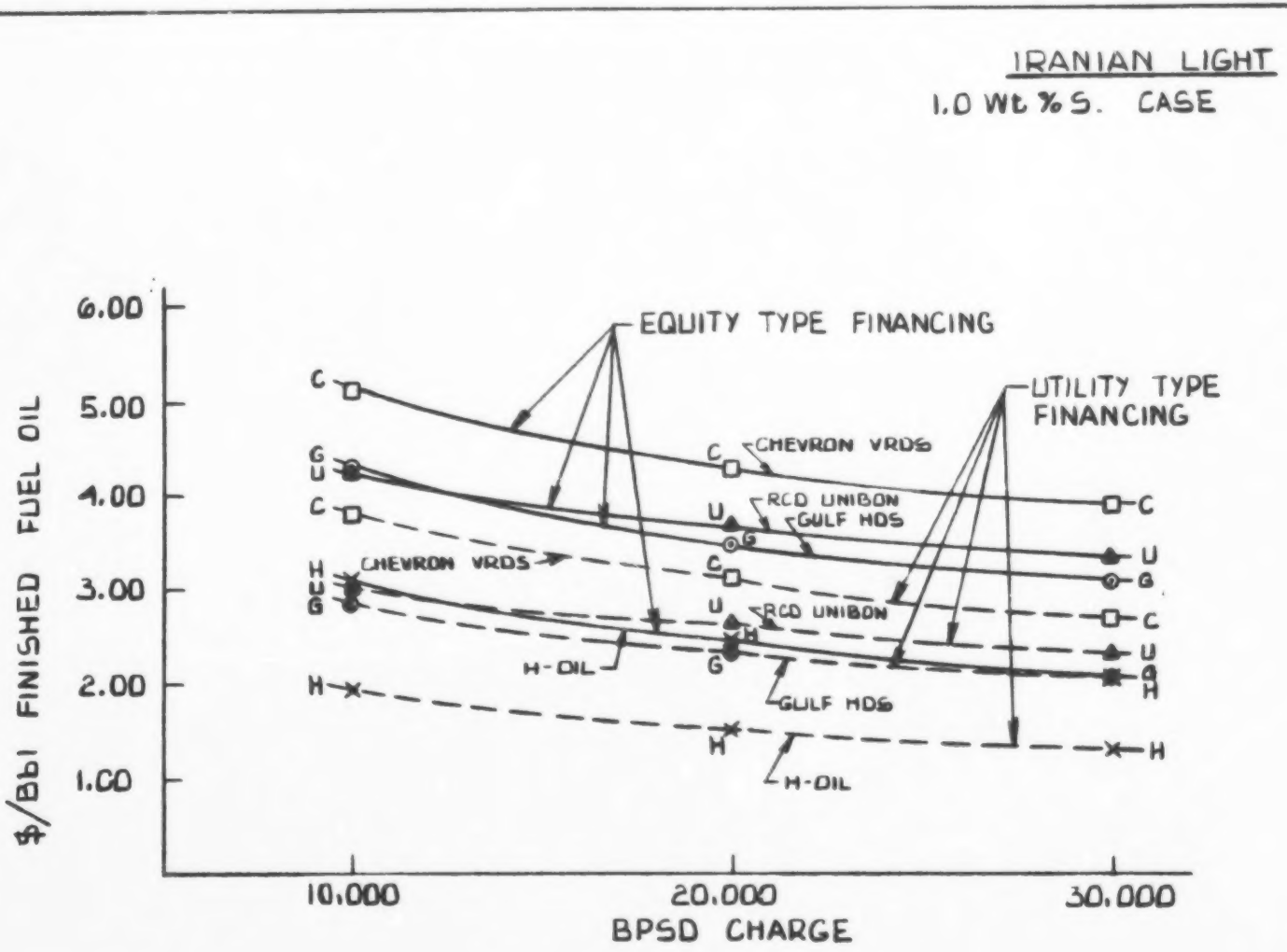


FIGURE 5-3-8

CLIENT ENVIRONMENT CANADA
PROJ. EVALUATION OF DESULPH. PROCESSES
COST OF DESULPHURIZATION
IRANIAN LIGHT RESIDUE
DRAWN N.M
CHKD PSK
APPR.
DATE
DATE REV.
NO
ECH./SCALE
ELEMENT
SUBDIV.
CONTR.
FORMAT A

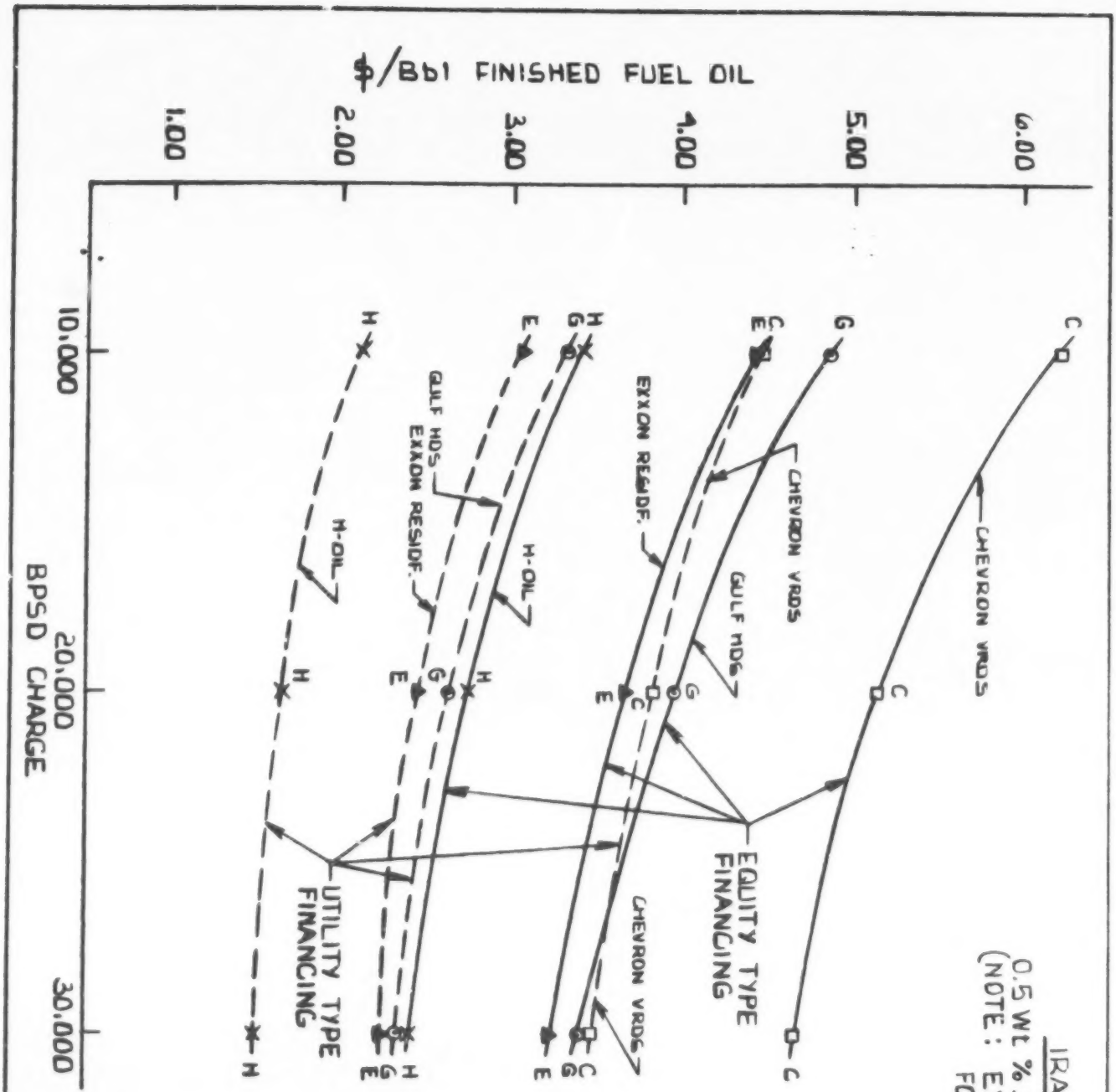


FIGURE 5-3-9

CLIENT ENVIRONMENT CANADA		FAIT/MADE N.M.	VERIFIE/CHKD PSL	DATE
PROJ. EVALUATION OF DESULPH. PROCESSES		APPR.	ECH./SCALE —	DATE REV.
COST OF DESULPHURIZATION IRANIAN LIGHT RESIDUE		CONTR. / SUBDIV.	ELEMENT	NO. / REV.

FORMAT "A"

CANADIAN LT-SOUR BLEND
1.0 Wt % S CASE +

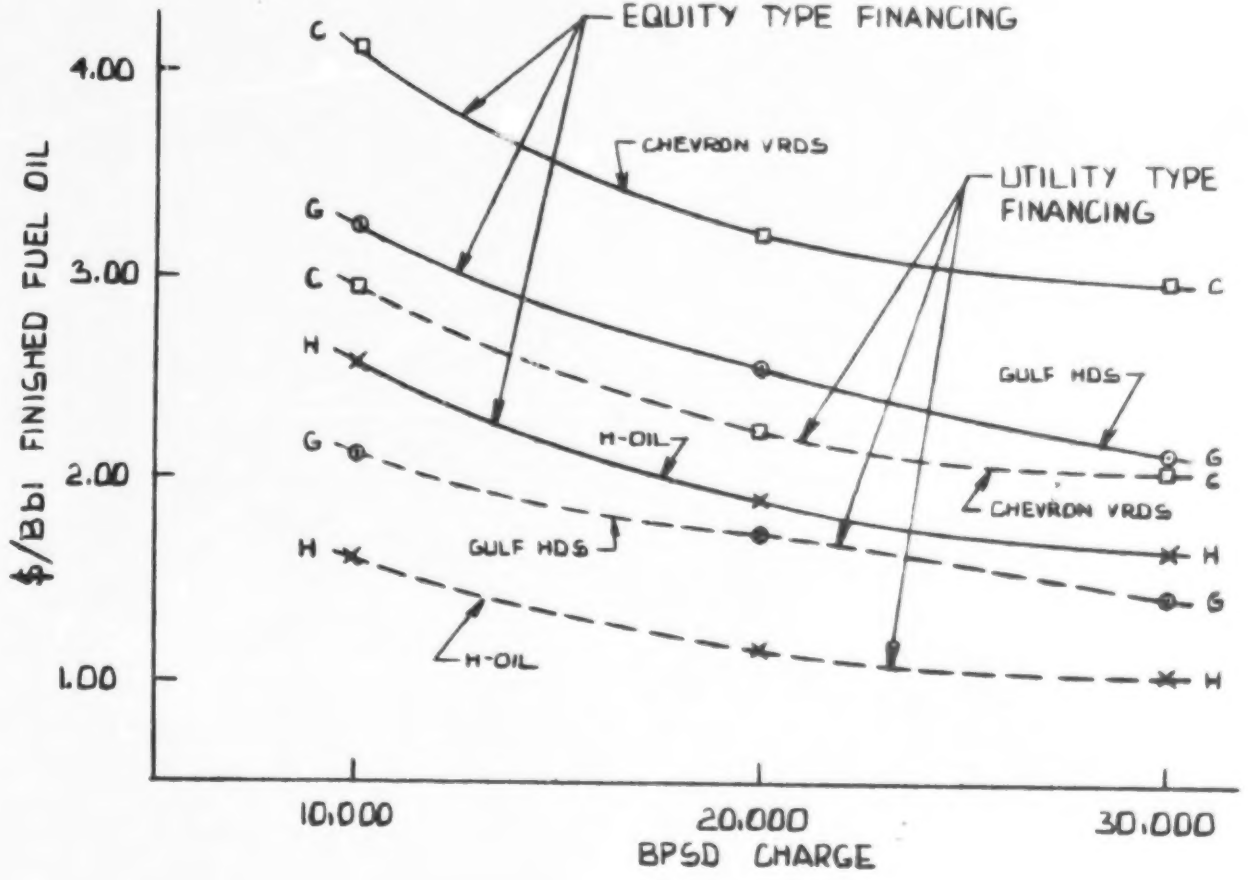
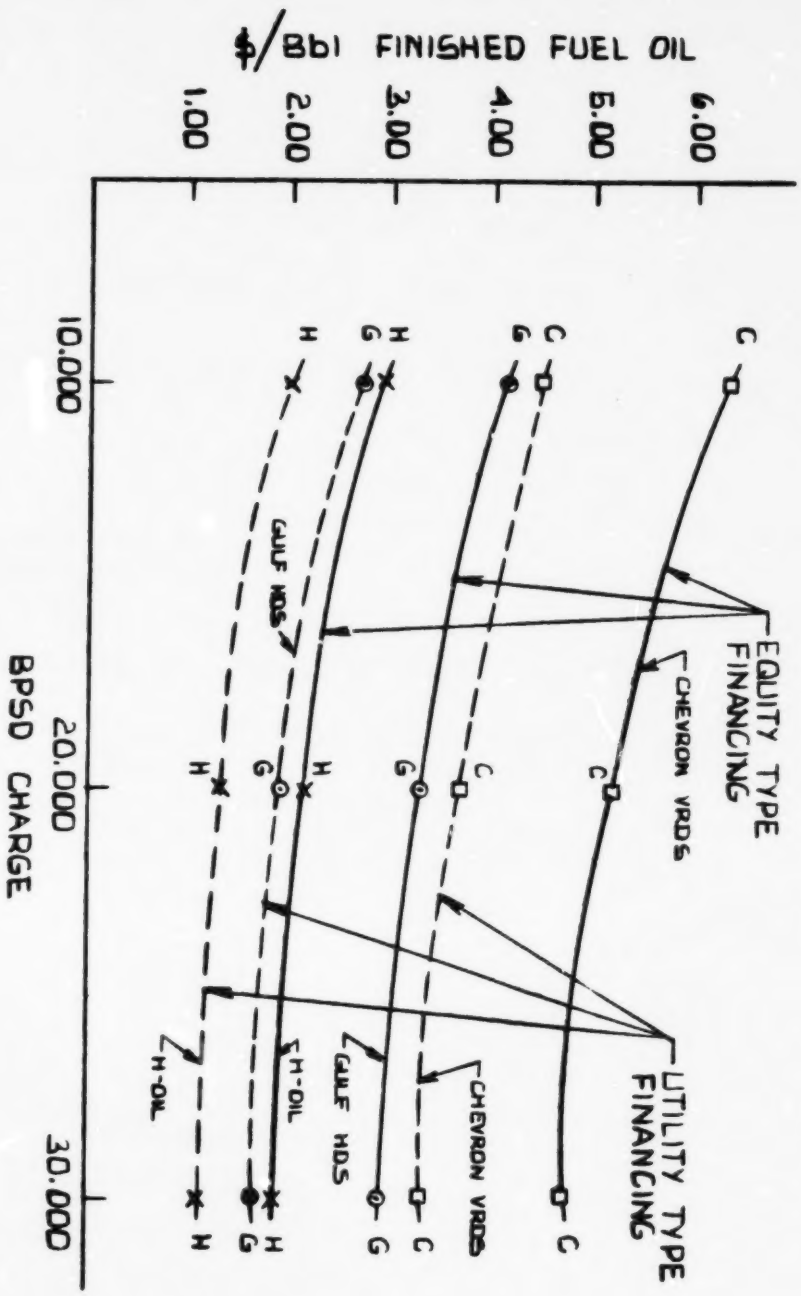


FIGURE 5-3-10

DATE		DATE REV	
VERIFIE/CHKD PSL		ECH./SCALE —	NO
FAIT/MADE N.M.		APPR	ELEMENT
CONTR.	SUBDIV.		

CLIENT ENVIRONMENT CANADA
PROJ. EVALUATION OF DESULPH. PROCESSES
COST OF DESULPHURIZATION
CANADIAN LT-SOUR BLEND RESID.

FORMAT "A"
3128-02/2



CANADIAN LT-SOURBLEND
0.5 Wt % S. CASE

FIGURE 5-3-11

CLIENT ENVIRONMENT CANADA	FAIT/MADE N.M.	VERIFIE/CHKD PSK	DATE
PROJ. EVALUATION OF DESULPH. PROCESSES	APPR.	ECH./SCALE —	DATE REV.
COST OF DESULPHURIZATION CANADIAN LT-SOURBLEND RESID	CONTR. SUBDIV.	ELEMENT	NO. REV

FORMAT "A"

5. TECHNICAL/ECONOMIC ASSESSMENT (cont'd)

5.4 Conclusions

The following generalized conclusions can be drawn from this study:

1) The economics of direct desulphurization is highly dependent on the metals (nickel and vanadium) content of the feed as well as the degree of desulphurization required.

2) The cost of desulphurization is highly sensitive to the capital charges and the costs of hydrogen and the HDS catalyst. There is no significant difference between the fixed and variable operating costs in the processes considered.

3) For direct desulphurization of vacuum residua with high levels of metals content (over 250ppm), H-Oil is the only viable process of all the processes considered.

4) H-Oil seems to have a significant cost advantage over the fixed bed processes. The H-Oil catalyst exhibits much higher hydrocracking activity leading to higher hydrogen consumption.

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APPENDIX A

BASIS FOR CALCULATING PRODUCTION COSTS

a) Fixed Operating Costs

Operating Personnel	- $15,330/year
Supervisory/Technical Personnel	- 25% of Operating Labour
Maintenance	- 3% of IBL Investment 1.5% of OBL Investment
General, Sales, Administrative	- 1.5% of Total (IBL + OBL) Investment
Taxes	- 1.0% of Total (IBL + OBL) Investment
Insurance	- 1.0% of Total (IBL + OBL) Investment

b) Variable Operating Costs

Electricity	- 1.2¢/KWH
Fuel	- $1.54/MMBTU
Steam* (50 psig) (450 psig)	- $2.60/Mlb. $2.75/Mlb.
Cooling Water*	- 6¢/1000 USg
Boiled Feed Water*	- $0.32/1000 USg

c) Byproduct Credits

Sulphur	- $27/LT
Unstabilized Naphtha	- $11/Bbl

d) Capital Charges†

		UTILITY FINANCING	EQUITY FINANCING
Depreciation	IBL		6% per year
	OBL		4% per year
Interest	IBL	- 10.3% per year	15% per year
	OBL	- 10.3% per year	15% per year
	WC**	- 10% per year	10% per year

Notes:

IBL = Inside Battery Limits
OBL = Outside Battery Limits
WC = Working Capital
HDS Catalyst Costs Increased by 5% for F.O.B. Edmonton

*include capital charges
**initial catalyst charge capitalized
†cost of land and royalties excluded

APPENDIX B

LICENSOR DATA

The attached tables outline the yield investment and utility data provided by the process licensors. In interpreting the data the following should be noted.

1. Except for H-Oil, none of the processes could handle Tia Juana residue on a direct desulphurization basis.

2. All process investment costs refer to the installed cost of process equipment within the battery limits but exclude the service facilities.

   Engineering Contractors' fees are, however, included. The investment costs are for U.S Gulf Coast location for first quarter 1977 unless otherwise indicated.

3. For Gulf HDS process, the following additional data were provided for 20,000 BPSD:

		Process Investment MM$	Catalyst Consumption MM$/Yr.
Iranian Light	1.0 Wt. %S	36.70	1.95
	0.5 Wt. %S	39.70	2.32
Canadian Light Sour Blend	1.0 Wt. %S	25.00	0.77
	0.5 Wt. %S	32.00	1.03

Two catalyst fills per year are required.

Gulf also provided the following hydrogen makeup (95 mole %) rates excluding mechanical losses.

	Iranian Light		Canadian Light Sour Blend	
Wt %S. in 375 F+ product	0.5	1.0	0.5	1.0
Hydrogen (95 mole %) makeup MMSCFD				
Start of Run	17.6	13.2	16.8	12.5
Average of Run	21.1	16.6	19.0	14.8
End of Run	25.9	21.3	22.0	18.0

4. For the Chevron VRDS process, the following additional data were provided for 5,000 BPSD.

		Process Investment MM$	Catalyst Consumption MM$/Yr.
Iranian Light	1.0 Wt. %S	9.80	1.40
	0.5 Wt. %S	12.00	1.60
Canadian Light Sour Blend	1.0 Wt. %S	8.80	0.70
	0.5 Wt. %S	13.00	1.20

The 5,000 BPSD cases involve single train - single reactor plant while the 30,000 BPSD cases are based on three parallel trains of two reactors each.

The Iranian Light residue cases would require catalyst replacement twice a year while the Canadian Light Sour Blend cases are based on once a year replacement.

5. H-Oil provided the following additional investment data for 20,000 BPSD.

		Process Investment MM$
Tia Juana	1.0 Wt. %S	24.00
	0.5 Wt. %S	38.00
Iranian Light	1.0 Wt. %S	18.00
	0.5 Wt. %S	23.00
Canadian Light Sour Blend	1.0 Wt. %S	14.00
	0.5 Wt. %S	20.00

HRI have indicated that a single train of two reactors could be adequate for 20,000 - 30,000 BPSD depending on the reactor fabrication and transportation aspects.

6. Adjustements to the overall utility balances have been made to account for the following terminal conditions assumed by the process licensors:

   a) Chevron utility estimates are based on feed and product temperatures of 450°F and 150°F respectively. Makeup H_2 is supplied at 250 psig and 120°F.

   b) The utility estimates provided by UOP are based on feed and product temperatures of 400°F and 120°F respectively. The makeup H_2 is supplied at 200 psig and 100°F. The steam consumption does not include fractionator stripping steam requirements.

   c) The Gulf data are based on feed and product temperatures of 200°F, makeup H_2 supply at 250 psig and 120°F. The steam consumption include the MEA regeneration requirements for the High Pressure Circuit.

   d) Exxon's assumed terminal conditions are: feed and product temperatures of 534°F and 560°F respectively. Makeup H_2 is supplied at 250 psig.

   e) The basis for H-Oil data are: feed and product at 650°F. Makeup H_2 at 400 psig and 100°F.

7. The process royalties are as follows for 30,000 BPSD/27,000 BPCD Units:

Gulf HDS	US$1,436,070
Chevron	1,971,000
UOP	2,714,000

GULF HDS DATA

	Iranian Light				Canadian Light Sour Blend			
i) Product Yields	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol %
H_2	(0.97)		(0.75)		(0.85)		(0.65)	
NH_3	0.20		0.14		0.53		0.38	
H_2S	2.72		2.20		3.11		2.59	
C_1	0.17		0.13		0.15		0.11	
C_2	0.18		0.14		0.16		0.11	
C_3	0.26		0.21		0.27		0.19	
C_4	0.21		0.17		0.21		0.15	
C_5 - 375°F Naphtha	2.31	3.04	1.80	2.32	2.62	3.47	1.85	2.44
375°F +	94.88	101.89	95.97	101.99	93.81	102.66	95.26	103.19
Wt. %S in 375 + Product	0.5		1.0		0.5		1.0	
ii) Product Properties								
C_5 - 375°F Naphtha								
°API			52.0					
Sulphur Wt. %	0.02		0.03		0.02		0.03	
375°F + Product								
°API	21.2		19.6		20.2		18.6	
Sulphur Wt. %	0.5		1.0		0.5		1.0	
Ni + V ppm	24		31.5		9.6		19.7	
iii) Investment and Utility Estimates (30,000 BPSD)								
Process Investment MM$	51.50		47.70		42.00		32.50	
Catalyst MM$/Yr.	2.53		2.00		1.35		1.00	
Power KW	10,190		9,475		9,375		8,956	
Cooling Water USgpm 20° ΔT	3,335		3,645		3,370		3,208	
Steam Mlb/hr.	25		19		31		25	
Boiler Feed Water USgpm	40		40		40		40	
Condensate USgpm	160		120		60		40	
Fuel MMBTU/hr.	100		100		100		100	

CHEVRON VRDS DATA

	Iranian Light				Canadian Light Sour Blend			
i) Product Yields	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol %
H_2 ··	(1.63)		(0.99)		(1.84)		(1.15)	
NH_3	0.27		0.19		0.50		0.35	
H_2S	2.98		2.45		3.08		2.55	
C_1	0.34		0.24		0.43		0.31	
C_2	0.41		0.30		0.52		0.38	
C_3	0.50		0.38		0.63		0.46	
C_4	0.38		0.27		0.48		0.35	
C_5 - 375°F Naphtha	2.07	2.70	1.47	1.90	2.73	3.6	1.90	2.5
375°F +	94.68	99.40	95.69	99.90	93.47	98.9	94.85	99.60
Wt. %S in 375°F + Product	0.5		1.0		0.5		1.0	
ii) Product Properties								
C_5 - 375°F								
°API	48.9		47.5		51.0		51.0	
Sulphur Wt. %	0.06		0.11		0.06		0.12	
375°F +								
°API	14.0		13.1		15.5		14.0	
Sulphur Wt. %	0.51		1.01		0.51		1.01	
iii) Investment and Utility Estimates (30,000 BPSD)								
Process Investment MM$	56.0		47.0		63.0		38.0	
Catalyst Initial Fill MM$	7.1		6.0		7.1		4.2	
Catalyst Replacement MM$/Yr.	9.4		8.1		7.1		4.2	
Power KW	5,790		3,930		6,930		4,430	
Cooling Water USgpm 20° Δ T	2,500		2,000		2,400		1,900	
Steam Mlb/hr.	(8)		(11)		(13)		(15)	
Boiler Feedwater USgpm	61		66		67		72	
Condensate USgpm	49		13		108		43	
Fuel MMBTU/hr.	52.5		52.5		52.5		52.5	

UOP RCD-UNIBON AND EXXON RESIDFINING DATA

i) Product Yields

	Iranian Light				Canadian Light Sour Blend	
	UOP		EXXON		UOP	
	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol %
Wt. %S in Product	1.0		1.0		1.0	
H_2	(0.83)				(1.36)	
NH_3	0.11		Exact Yields Not Provided		0.39	
H_2S	2.17				2.55	
C_1	0.05				0.08	
C_2	0.04				0.07	
C_3	0.05				0.09	
C_4	0.06				0.09	
C_5 - 375°F Naphtha	1.61	2.08			2.17	3.14
375°F +	96.75	100.22			95.92	100.00

ii) Product Properties

C_5 - 375 Naphtha			
°API	54.0	N/A	41.0
Sulphur Wt. %	N/A	N/A	$\leq$0.2
375°F + Product			
°API	15.8	N/A	13.0
Sulphur Wt. %	1.0	N/A	1.02

iii) Investment and Utility Estimates (30,000 BPSD)

Process Investment MM$	45.33	38.9*	56.0*
Catalyst MM$/Yr.	6.48	5.27	2.85
Power KW	6,025	6,340	6,025
Cooling Water USgpm, 20° ΔT	1,037	2,270	1,037
Steam Mlb/hr.	13	(80)	13
Boiler Feed Water USgpm	-	188	-
Condensate USgpm	18	42	18
Fuel MMBTU/hr.	45	42.5	45

* Second quarter 1977

4. H-OIL DATA

	Tia Juana				Iranian Light				Canadian Light Sour Blend			
	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol %	Wt. %	Vol
Wt. %S in 375°F⁻ Product	0.5		1.0		0.5		1.0		0.5		1.0	
i) **Product Yields**												
H_2	(1.9)		(1.2)		(1.4)		(1.1)		(1.6)		(1.1)	
$NH_3 - H_2S$	3.1		2.5		3.0		2.3		3.2		2.9	
$C_1 - C_3$	3.6		1.7		2.0		1.4		2.7		1.6	
C_4 - 375°F	10.3	15.0	6.2	9.0	7.0	9.7	5.4	7.5	9.7	13.9	5.9	
375°F⁺	84.9	92.1	90.8	96.4	89.4	95.3	92.0	96.6	86.0	93.5	90.7	96.!
ii) **Product Properties**												
C_4 - 375°F Naphtha												
°API	68		68		66		65		61		61	
Sulphur Wt. %	0.09		0.07		0.07		0.07		0.07		0.07	
375°F⁺ Product												
°API	18.2		15.0		20.1		17.8		19.2		16.0	
Sulphur Wt. %	0.51		1.00		0.5		1.00		0.51		1.00	
iii) **Investment and Utility Estimate (10,000 BPSD)**												
Process Investment MM$*	22.40		14.30		13.70		10.80		12.20		8.72	
Catalyst MM$/Yr.	1.42		0.63		0.56		0.47		0.82		0.55	
Power KW	4,142		3,308		3,262		2,829		3,151		2,547	
Cooling Water USgpm, 20° Δ T	235		178		191		162		157		115	
Steam Mlb/hr.	3.553		3.553		3.553		3.553		2.50		2.50	
Condensate USgpm	20		20		20		20		16		16	
Fuel MMBTU/hr	24		16		18		13		8		10	

*includes initial catalyst load

APPENDIX C

APPROACH TO CAPITAL COST ESTIMATING

The capital costs presented in this report were derived from the individual process unit equipment lists (Section 5) established for the base cases using inhouse data supplemented by the budget type estimates from shop fabricators. Material and labour modules were developed for the Edmonton location to arrive at the installed cost of process equipment.

To these "bare" costs, adjustments were made to include:

- Site clearance, maintenance gravel roads and access roads between the process units.
- Tie-in of utilities such as steam, cooling water, etc., which were assumed available from the existing refinery complex.
- Hydrogen booster compressor to supply the HDS unit.
- Auxiliaries such as power supply to MCC, firewater loops with hydrants and monitors, oily water sewer system, potable water and emergency wash facilities, hydrocarbon blowdown systems, flare header and connected to the refinery main system and control room with a small testing laboratory.
- Engineering contractors' fees for Engineering Procurement-Construction management.

A comparison of the total investment estimate thus derived with the estimates provided by the licensors for U.S.A Gulf Coast location showed the latter consistently below the Edmonton base costs by at least 30%.

The higher capital costs estimated in this study can be explained by about 5-10% allowance for above mentioned items which have not been included in the licensor supplied estimates as well as 20-25% differential in installed costs between Edmonton and U.S.A Gulf Coast location. The latter are due to lower labour productivity and higher delivered costs of equipment at Edmonton compared to the Gulf Coast which is in agreement with a recent survey (12) carried out by Imperial Oil Limited as well as a study carried out for the Ministry of Trade and Commerce (13).